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PARAMETRIC STUDY OF POTENTIAL EARLY COMMERCIAL MHD POWER PLANTS

C. H. Marston, et al
General Electric
Space Division

February 1980

Prepared for
NATIONAL AERONAUTICS AND SPACE ADMINISTRATION
Lewis Research Center
Under Contract DEN 3-52

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for
U.S. DEPARTMENT OF ENERGY
Office of Magnetohydrodynamics
Assistant Secretary for Fossil Energy

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16. Abstract

A parametric study was performed to determine performance and cost of "moderate technology" coal-fired Open Cycle MHD/steam power plant designs which can be expected to require a shorter development time and have a lower development cost than previously-considered "mature" OCMHD/Steam plants. Three base cases were considered: Base Case 1 had an indirectly-fired high temperature air heater (HTAH) subsystem delivering air at 2700 F; fired by a state-of-the-art atmospheric pressure gasifier; Base Case 2 had an indirectly-fired HTAH subsystem delivering air at 3000 F; in Base Case 3 the HTAH subsystem was deleted and oxygen enrichment was used to obtain requisite MHD combustion temperature.

Coal pile to bus bar efficiencies in Base Case 1 ranged from 41.4% to 42.9%, and cost of electricity (COE) was highest of the three base cases. For Base Case 2 the efficiency range was 42.0% to 45.6%, and COE was lowest. For Base Case 3 the efficiency range was 42.9% to 44.4%, and COE was intermediate. The best parametric cases in Base Cases 2 and 3 are recommended for conceptual design. Eventual choice between these approaches is dependent on further evaluation of the tradeoffs among HTAH development risk, O_2 plant integration and further refinements of comparative costs.

17. Key Words (Suggested by Author(s))	18. Distribution Statement
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Table i-1.

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FOREWORD

The work described in this report is a part of the Parametric Study of Prospective Early Commercial MHD Power Plants (PSPEC) sponsored by the MHD Division of the U.S. Department of Energy (DOE) and directed by the Lewis Research Center of the National Aeronautics and Space Administration (NASA LeRC).

This General Electric Contractor report covers Task I of the study, which was parametric analysis. The prime contract was with General Electric's Space Sciences Laboratory, MHD Programs Section. Members of the technical staff of that and the following organizations participated in this study.

Bechtel National, Inc.:	Balance of Plant and Costing
Foster Wheeler Development Corp.:	Chemically Active Fluidized Bed Combustor
General Electric Energy Systems Products Department:	Superconducting Magnet Evaluation
Hooker Chemical Co.:	Seed Reprocessing Evaluation

This General Electric contractor report covers one of two parallel studies. AVCO Everett Research Laboratory, Inc. was the other prime contractor.

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NOTE ON STEAM PLANT CONFIGURATION, PERFORMANCE AND COST DATA

A 1977 consent decree places certain restrictions on the General Electric Company regarding the furnishing of information on steam turbine-generator sets. Therefore, the steam power plant configurations and performance for this study were determined solely on the basis of published information*. They are representative of a modern steam plant, but not based on specific hardware and not necessarily consistent with any guaranteed heat rate. Cost estimation was done by the Architect and Engineer Subcontractor, Bechtel National, Inc., on the basis of their own data. Cost and performance data presented herein are for study purposes only. Nonetheless we believe that the data are generally accurate enough for the intended purpose of comparing cost and performance among the MHD/steam plants considered.

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*Spencer, R. C., Cotton, K. C. and Cannon, C. N., ASME Journal of Engineering for Power, Oct. 1963.

SECTION 1

INTRODUCTION

SECTION 1

INTRODUCTION

1.1 BACKGROUND

Results for Open Cycle MHD from The Energy Conversion Alternatives Study (ECAS)[1,2,3,4,5] did much to encourage a development program leading to large scale, commercial, coal fired, OCMHD topping/steam bottoming power plants. Subsequent studies of the Engineering Test Facility (ETF)[6,7,8] are providing conceptual designs of an intermediate size plant to demonstrate engineering feasibility and technological readiness and to provide data for scale-up to a commercial plant.

These studies are typical of the two approaches needed to identify the most promising MHD power plant systems from the standpoints of fuel efficiency and cost effectiveness and to provide reasonable assurance that components can be developed with the requisite performance, durability and reliability. These two approaches are:

1. Top Down - For an array of possible systems, examined under a uniform set of basic assumptions and ground rules, what performance and costs can be expected, and what is required of each system's components and subsystems? How can system performance be improved?

2. Bottom Up - How would one design an MHD plant which could be built with present or foreseeable materials, tooling and construction techniques, always keeping in mind the effects on other parts of the system of a particular design choice within a component or subsystem?

The first approach, typified by ECAS, provides an identification of systems with promise, of specific problems or limitations which must be overcome in otherwise promising systems, and of apparently fundamental limitations which may rule out some systems. The identification of an MHD power plant as "promising" must of course be made in a broader context than that looked at in a particular study. OCMHD must compete with existing systems which have the twin advantage of many years of development and an established performance record as well as with other proposed systems ranging from modest advances in existing steam, gas turbine and combustion technology to fusion reactors and satellite solar power plants. The second approach, typified by ETF, provides insight into system interactions, specific design problems, and deficiencies in data, materials or techniques which are not brought out in a top down study. The results of both must be subject to continuing review in the light of new insights, new developments and a better appreciation of the impact of assumptions and ground rules.

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1 References are listed in Section 7

This Parametric Study of Potential Early Commercial MHD Power Plants (PSPEC) is an appropriate follow-up to ECAS and ETF. As a new "top down" study it had the benefit of:

1. A better appreciation of the impact of parametric choices and their interrelation.
2. A better appreciation of system and component design constraints plus design data at ETF size for scale-up.
3. Improved analytical tools for system and component performance prediction.

1.2 SCOPE

The focus of PSPEC Task I has been on parametric examination of performance and cost of moderate technology coal fired OCMHD/steam power plant designs which can be expected to require a shorter development time and have a lower development cost than the directly fired air heater system which was the culmination of the ECAS effort.

Three base cases were considered in PSPEC as indicated in Table 1.2-1. For each of these base cases a reference case and a series of parametric cases were defined. For most of the parametric cases a single major parameter was varied (e.g., combustor type, coal type, magnetic field), but some other variables were also adjusted (e.g., combustor pressure, radiant furnace duty) for optimum performance. More complete specifications for the cases are given in Section 1.5, system results are presented in Section 2 and details of subsystems are in Section 3.

1.3 DEFINITIONS AND REFERENCE CONDITIONS

The definitions in Table 1.3-1 are used throughout the study.

Table 1.3-1. Definitions

FLOW TRAIN THERMAL POWER:	COAL HHV PLUS COAL, SEED AND OXIDIZER SENSIBLE HEAT TO MAIN FLOW TRAIN
OXYGEN ENRICHMENT:	KG PURE O_2 ADDED PER 100 KG OF MOIST AIR
REFERENCE ENTHALPY:	HEAT OF FORMATION AT 298K (77F) OF FUEL, SEED AND OXIDIZER TO THE COMBUSTOR
FUEL MOISTURE RATIO (FMR):	KG WATER PER 100 KG DRY COAL

Table 1.2-1. Case Overview

BASE CASE 1

COMMON: 2700 F AIR PREHEAT, ATMOSPHERIC PRESSURE GASIFIER FOR HTAH

ITEM	REFERENCE	VARIATION
COAL	MR	I6
OXIDIZER	AIR + 10% O_2	AIR
MHD COMBUSTOR	2 STAGE CYCLONE	1 STAGE VORTEX
B FIELD	(6-5) TESLA	(7-6) TESLA

BASE CASE 2

COMMON: 3000 F AIR PREHEAT

ITEM	REFERENCE	VARIATION(S)
SIZE (NOMINAL)	1200 MWe	900 MWe, 600 MWe
COAL	MR	I6
AIR HTR COMBUSTOR	2-STAGE, PRESS	1-STAGE, CAPFB,* ATM PRESS
AIR HEATER INLET TEMP	600 F	1300 F
MHD COMBUSTOR	1-STAGE	S^3PMB**, 2-STAGE
B FIELD	(6-5) TESLA	(8-7) T, E_Y CONSTANT
MHD FLOW	SUBSONIC	SUPERSONIC
MHD FLOW TRAIN	SINGLE	DUAL

BASE CASE 3

COMMON: NO HTAH, AIR + 40% O_2

ITEM	REFERENCE	VARIATION
COAL	MR	I6
PREHEAT	1300 F	1100 F
MHD COMBUSTOR	1-STAGE	2-STAGE
B FIELD	(6-5) TESLA	(8-7) TESLA

*CHEMICALLY ACTIVE PRESSURIZED FLUIDIZED BED

**SPLIT STREAM SLAGGING PRESSURIZED MOVING BED GASIFIER

Flow train thermal power as defined in Table 1.3-1 is useful in calculation of channel performance because it fixes channel size, independent of whether the air heaters, if any, are directly or indirectly fired. The definition does not include losses in an indirectly fired air heater system but these are, of course, included in calculations of coal-pile-to-bus-bar efficiency.

Oxygen enrichment can be expressed in several ways but the definition shown in Table 1.3-1 was the most useful one for our equilibrium analysis. Figure 1.3-1 provides conversion to O_2% by weight and volume and to N_2/O_2 ratio.

Reference enthalpy is the standard definition, consistent with JANAF thermochemical properties. For purposes of computing compressor work and heat rejection, the ambient temperature was assumed to be 59 F, however, all energy and combustion calculations are on the basis of the 77 F (25 C) standard reference temperature.

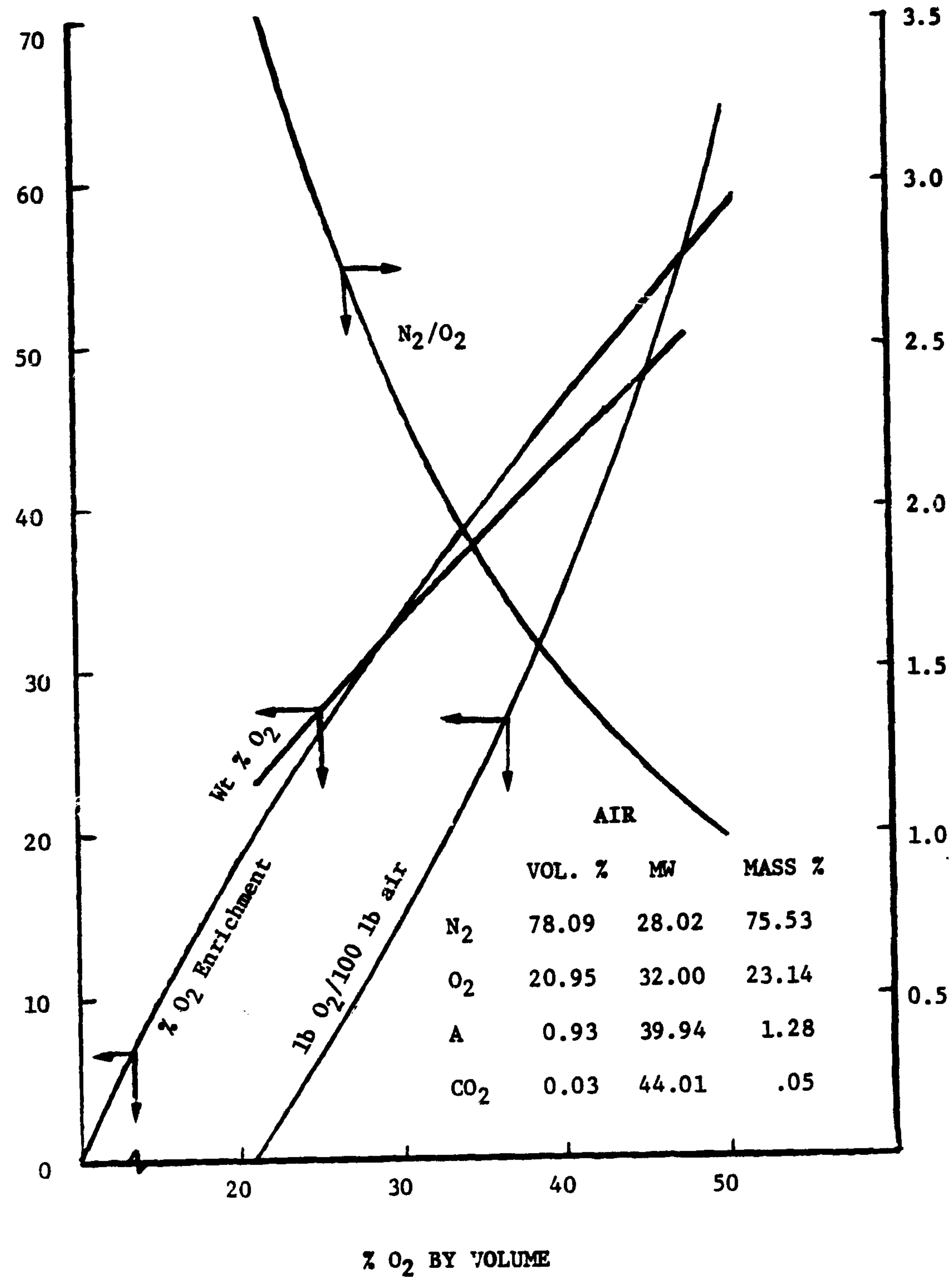
1.4 COAL AND AIR COMPOSITION

Montana Rosebud (MR) and Illinois #6 (I6) coals were specified for PSPEC. Montana Rosebud was used for all three reference cases. Composition for these coals is given in Tables 1.4-1 and 1.4-2 and is consistent with specifications used for ETF. Some additional properties shown in the tables have been assumed on the basis of data in Reference 9 and 10. The 'as received' fuel moisture ratios (FMR = Kg Fuel H_2O/100 Kg Dry Coal) were 29.37 for MR and 9.77 for I6. Except for some gasifier cases, coal to the main combustor was dried to a FMR of 5 and 2 for MR and I6 respectively. These fuel moisture levels were assumed to be the nominal water content limits for conventional coal drying without loss of coal heating value from coal devolatilization.

Air composition and properties are given in Table 1.4-3. Sample Mollier Charts for products of combustion are included in Section 1.6.

Table 1.4-3. Air Composition and Properties

Reference Temperature		Ambient Pressure	Humidity
77 F		14.7 psi	0.65 Kg H_2O per 100 Kg Dry Air
	Component	Wt%	
	O_2	23.49	
Dry Air	CO_2	0.046	
Composition	A	1.286	
	N_2	75.519	



AIR

	VOL. %	MW	MASS %
N_2	78.09	28.02	75.53
O_2	20.95	32.00	23.14
A	0.93	39.94	1.28
CO_2	0.03	44.01	.05

Figure 1.3-1. Oxygen Enrichment Conversion Chart

Table 1.4-1. Coal Analysis, Montana Rosebud Coal

SAMPLE DESCRIPTION:	Montana Rosebud Coal (Specified for PSPEC)	Subbit B

TOTAL MOISTURE (AS REC'D)	% 22.7
AIR DRY MOISTURE LOSS	% 12.7
EQUILIBRIUM MOISTURE	%

PROXIMATE ANALYSIS

	As Rec'd	Dry
% Fixed Carbon	39.2	50.7
% Volatile Matter	29.4	38.0
% Ash	8.7	11.3
% Moisture	22.7	--
TOTAL	100.0	100.0

ULTIMATE ANALYSIS

	d.a.f	As Rec'd	Dry
% C	76.01	52.10	67.4
% H	5.08	3.48	4.5
% O	16.52	11.32	14.65
% N	1.15	0.79	1.02
% S	1.24	0.85	1.10
% Ash	--	8.76	11.33
% Moisture	--	22.70	--
% Cl	--	--	--
TOTAL	100.0	100.0	100.0

SULFUR COMPOUNDS

	As Rec'd	Dry
Sulfate		
Pyritic		
Organic		

ASH FUSION TEMPERATURE, °F

	Red Atm.	Oxid. Atm.
Initial Deform.	2190	
Soft Temp. Sph.	2230	
Soft Temp. Hem.		
Fluid Temp.	2280	

REACTIVITY

Reactivity Index	
Energy of Activation	$\frac{cal^{\circ}C}{g\ mol}$

H.H.V.	As Rec'd	8,936
BTU/lb.	Dry	11,560

Specific Gravity	

HARDGROVE INDEX	50
FREE SWELLING INDEX	1

DEPOSITS OR ASHES

% WT OF DEPOSIT

SiO_2 37.6	P_2O_5 0.4	K_2O 0.5
Al_2O_3 17.3	CaO 11.0	SO_3 17.5
Fe_2O_3 5.1	MgO 4.0	Cl 10^{-6}
TiO_2 0.7	Na_2O 3.1	Other 2.8

Table 1.4-2. Coal Analysis, Illinois #6 Coal

SAMPLE DESCRIPTION:	Illinois #6 Coal (Specified for PSPEC)	hvCb

TOTAL MOISTURE (AS REC'D)	% 8.9
AIR DRY MOISTURE LOSS	% 6.9
EQUILIBRIUM MOISTURE	% ___

PROXIMATE ANALYSIS

	As Rec'd	Dry
% Fixed Carbon	41.7	45.8
% Volatile Matter	38.0	41.7
% Ash	11.4	12.5
% Moisture	8.9	--
TOTAL	100.0	100.0

ULTIMATE ANALYSIS

	d. a. f	As Rec'd	Dry
% C	78.29	62.40	68.50
% H	5.53	4.41	4.84
% O	10.53	8.39	9.21
% N	1.51	1.20	1.32
% S	4.14	3.30	3.62
% Ash	--	11.40	12.51
% Moisture	--	8.90	--
% Cℓ	--	--	--
TOTAL	100.0	100.0	100.0

SULFUR COMPOUNDS

	As Rec'd	Dry
Sulfate		
Pyritic		
Organic		

ASH FUSION TEMPERATURE, °F

	Red Atm.	Oxid. Atm.
Initial Deform.	1960	
Soft Temp. Sph.	2030	
Soft Temp. Hem.		
Fluid Temp.	2260	

REACTIVITY

Reactivity Index	
Energy of Activation	$\frac{cal\,^{\circ}C}{g\ mol}$

H.H.V.		
BTU/lb.	As Rec'd	11,269
	Dry	12,370

Specific Gravity	

HARDGROVE INDEX	60
FREE SWELLING INDEX	6.0

DEPOSITS OR ASHES

% WT OF DEPOSIT

SiO_2	41.4	P_2O_5	0.12	K_2O	2.1
Al_2O_3	19.3	CaO	5.4	SO_3	7.5
Fe_2O_3	22.3	MgO	0.30	Cl	10^{-6}
TiO_2	0.9	Na_2O	0.6	Other	0.8

1.5 RATIONALE FOR CASE SELECTION

Tables 1.5-1, 1.5-2 and 1.5-3 show, for Base Cases 1, 2 and 3, respectively, the major features of the reference cases, and parametric variations studied for each, along with the principal design parameters. Note that neither operating pressure nor loading parameter are given for the MHD generator. These are determined as a result of generator/combustor optimization within the constraints specified.

Base Case 1 was restricted to a state-of-the-art gasifier and a High Temperature Air Heater (HTAH) system delivering 2700 F and was important to demonstrate what could be done with such a system. However, it was clear at the outset that efficiency would be relatively low. Parametric variations were therefore limited to changes in coal type, oxidizer, MHD combustor and magnetic field strength. The latter was restricted to a maximum of 7 Tesla because that was judged the maximum feasible with current niobium-titanium superconductor technology.

Base Case 2, with an HTAH system delivering 3000 F air, was expected to yield the best efficiency and was therefore the Base Case for which several different system arrangements, sizes, and MHD generator configurations were considered, as well as variations analogous to those in Base Case 1. Some combinations of two or more variations were considered, but the cases were constructed so that the effect of any single variation could be directly evaluated.

In keeping with the somewhat more advanced technology implicit in Base Case 2, several types of variations were also considered to evaluate the potential gains and/or costs to be expected in areas requiring substantial development. For example, cesium seed was assumed in Case 2.5. Sufficient raw material in the form of pollucite and other cesium ores exist but, in the absence of any significant market, cesium salts are not marketed in quantity. Similarly, an 8T magnetic field was assumed for Case 2.7, recognizing that structural requirements are much more severe and that niobium-tin or some as yet undiscovered superconductor would be required. The required case in which NASA, LeRC specified MHD generator performance was also included in Base Case 2. This was done in close cooperation with them and subsequent recalculation using NASA specifications with the GE codes duplicated their MHD generator output within 0.2%.

Because most of the exploratory work was done in Base Case 2, Base Case 3, with O_2 enrichment but with no HTAH, was limited to a few basic variations, as indicated in Table 1.5-3.

Two areas of plant design which impact overall system efficiency were held essentially constant in this study. These areas are the heat recovery/seed recovery (HRSR) subsystem and the steam cycle. The HRSR subsystem is a moderate-slag carryover, indirectly-fired HTAH concept similar to that proposed by Combustion Engineering for the AVCO/C-E ETF[7]. The steam subsystem is a supercritical 1000 F/1000 F/3500 psia cycle, with a feedwater heater train that incorporates a low temperature economizer and the MHD channel cooling.

1.6 OVERVIEW OF ANALYTICAL PROCEDURE

For each reference case and for several alternative system configurations in Base Case 2, an interactive systems code, designated OCSYS, was used to generate a complete system balance.

Table 1.5-1. Case List Base Case 1

	1.0	1.1	1.2	1.3	1.4	1.4a
Primary Change from Reference Case	Ref.	I6	Air	7T	Single Stage	Slag Rejection
Power Input, MWt*	2800	—	–	—	–	–
Coal Type	MR	I6	MR	—		
Oxidizer	Air + 10% O_2	– –	Air	Air + 10% O_2		
Seed, % Metal in Total Flow	1.3%	—	1% K	1.3%	—	—
Air Heater	IF	—	–			
Heater Combustor	Gasifier	—	—	—	—	—
Atmospheric/Pressurized	Atm					
Gasifier	Wellman					
Air Temperature, F	2700		—	—	—	
Sulfur Cleanup	Spray					—
Main Combustor	2-Stage Cyc				1-Stage Cyc	—
Slag Rejection, %	85					70
MHD Generator	Faraday	–		—	—	—
Length, m	25					—
Initial Mach Number	0.8					
Final Mach Number	1.0					
Magnetic Field, T	6-5					—
Diffuser Recovery	0.6	—				
Seed Reprocessing	No					

*Thermal power to 'HD combustor

Table 1.5-2a. Case List Base Case 2

	2.0	2.0s	2.0a	2.0b	2.1	2.2	2.2a	2.4	2.4a	2.5	2.6	2.7
Primary Change from Reference Case*	Ref.	Slag	S^3PMB + Coal	S^3PMB + SPMB	I6	2-Stage Cyc. Hot Bottom HTAH	2-Stage Cyclone	NASA Spec.	GE Recalc.	Cs Seed	Super Sonic	8 T
Power Input, MWt**	2800	—	—	—	—	—	—	—	—	—	—	—
Coal Type	MR	—	—	—	I6	MF	—	—	—	—	—	—
Oxidizer	Air	—	—	—	—	—	—	—	—	—	—	—
Seed	1% K	—	—	—	—	—	—	—	—	1% Cs	1% K	—
Air Heater (Indirectly Fired)	IF	—	—	—	—	—	—	—	—	—	—	—
Heater Combustor	2-Stage Cyclone	—	Gas-Phase Burner	—	2-Stage Cyclone	—	—	—	—	—	—	—
Atmospheric/ Pressurized	Press.	—	—	—	—	—	—	—	—	—	—	—
Gasifier	No	—	—	—	—	—	—	—	—	—	—	—
Air Temperature, F	3000	—	—	—	—	—	—	—	—	—	—	—
Sulfur Cleanup	Spray	—	—	—	—	—	—	—	—	—	—	—
Main Combustor	1-Stage Cyclone	—	S^3PMB + Coal	S^3PMB + SPMB	1-Stage Cyclone	2-Stage Cyclone	—	—	—	1-Stage Cyclone	—	—
Slag Rejection	85	70	91	> 99	70	85	—	—	—	—	—	—
MHD Generator	Faraday	—	—	—	—	—	—	—	—	—	—	—
Length, m	25	—	—	—	—	—	—	20	20	25	—	—
Inital Mach No.	0.8	—	—	—	—	—	—	0.9	0.8	—	1.2	0.3
Final Mach No.	1.0	—	—	—	—	—	—	<1.0	1.0	1.0	—	—
Magnetic Field, T	6-5	—	—	—	—	—	—	See Note	—	6-5	—	8-7
Diffuser Recovery	0.6	—	—	—	—	—	—	—	—	—	—	—
Seed Reprocessing	Formate	—	—	—	—	—	—	—	—	—	—	—

*For cases with letter modification variations are with respect to that numbered case.

Note on magnetic field: Cases 2.4, 2.4a, 2.16 and 2.16b are based on magnetic field prescribed to maintain E_y constant at 4 kV/m. B_{max} = 7.5 T

**Thermal power to MHD combustor

Table 1.5-2b. Case List Base Case 2 (Cont'd)

Primary Change from Reference Case*	2.10 1500 MWt	2.11 2000 MWt	2.11a Slag Rej.	2.12 Atm Air Htr Comb	2.15 Dual Flow Train	2.16 Hot Bottom HTAH, 4 kV/m Transverse Electric Field	2.16a Hot Bottom HTAH, 6-5 T	2.16b Cold Bottom HTAH, 4 kV/m Transverse Electric Field	2.17 CAPFB Air Htr Comb.	2.18 1300 F Air and Flue Gas to Preheat Combustor
Power Input, MWt **	1500	2000	—	2800	—	—	—	—	—	—
Coal Type	MR	—	—	—	—	—	—	—	—	—
Oxidizer	Air	—	—	—	—	—	—	—	—	—
Seed	1% K	—	—	—	—	—	—	—	—	—
Air Heater (Indirectly Fired)	IF	—	—	—	—	—	—	—	—	—
Heater Combustor	2-Stage Cyclone	—	—	1-Stage Air Cooled Cyc	2-Stage Cyclone	—	—	1-Stage Air Cooled Cyc	CAPFB/ Gas Phase Burners	1-Stage Air Cooled
Atmospheric/ Pressurized	Press	—	—	Atm	Press	—	—	—	—	—
Gasifier	No	—	—	—	—	—	—	—	—	—
Air Temperature, F	3000	—	—	—	—	—	—	—	—	—
Sulfur Cleanup	Spray	—	—	—	—	—	—	—	CAPFB	Spray
Main Combustor	1-Stage Cyclone	—	—	—	—	—	—	—	—	—
Slag Rejection	85	—	70	85	—	—	—	—	—	—
MHD Generator	F	—	—	—	—	—	—	—	—	—
Length, m	15	20	25	—	—	—	—	—	—	—
Initial Mach No.	0.8	—	—	—	—	—	—	—	—	—
Final Mach No.	1.0	—	—	—	—	—	—	—	—	—
Magnetic Field, T	6-5	—	—	—	—	See Note	6-5	See Note	6-5	—
Diffuser Recovery	0.6	—	—	—	—	—	—	—	—	—
Seed Reprocessing	Formate	—	—	—	—	—	—	—	—	—

*For cases with letter modification variations are with respect to that numbered case.

Note on magnetic field: Cases 2.4, 2.4a, 2.16 and 2.16b are based on magnetic field prescribed to maintain E_y constant at 4 kV/m. $B_{max} = 7.5$ T

**Thermal power to MHD combustor

Table 1.5-3. Case List Base Case 3

Primary Change from Reference Case	3.0 Ref.	3.1 I6	3.2 2-Stage Cyc.	3.4 1100 F Air	3.5 ST
Power Input, MWt*	2800	—	—	—	—
Coal Type	NR	I6	NR	—	—
Oxidizer	Air + 40% O_2	—	—	—	—
Seed	1.7% K	—	—	—	—
Air Heater	Recup only	—	—	—	—
Heater Combustor	NA			—	—
Atmospheric/Pressurized	NA	—			—
Gasifier	NA				
Air Temperature, F	1300		—	1100	1300
Sulfur Cleanup	NA	—	—	—	—
Main Combustor	1-Stage Cyc.		2-Stage Cyc.	1-Stage Cyc.	—
Slag Rejection, %	70		85	70	—
MHD Generator					
Length, m	25				—
Initial Mach Number	0.8				
Final Mach Number	1.0				
Magnetic Field, T	6-5				—
Diffuser Recovery	0.6	—	—	—	—
Seed Reprocessing	Formate		—	—	—

*Thermal power to MHD combustor

Other cases were then treated as perturbations from one of these detailed runs by adjusting the energy flows.

Inputs and outputs for code OCSYS are summarized in Table 1.6-1. The combustor and MHD generator analyses were treated separately as detailed in Section 3. In brief, the optimum net MHD generator power* and combustor operating pressure were determined for specified generator exit pressure, magnetic field distribution (or electrical stress), Hall voltage limit and entrance and exit Mach number. As described in Section 3.3.1, scaling equations for combustor heat loss as a function of pressure, temperature and mass flow rate were used to obviate the need for iteration between generator and combustor analysis.

Table 1.6-1. OCSYS System Code I/O

INPUTS

- MOLLIER CHARTS FOR PLASMA PROPERTIES (AS GENERATED FOR MHD GENERATOR ANALYSIS PLUS SUPPLEMENTS FOR LOW TEMPERATURE AND FOR HTAH SUBSYSTEM)
- INPUT MASS FLOW RATE AND ENERGY TO THE MHD COMBUSTOR
- MHD GENERATOR INLET STATE AND OUTPUT POWER
- THERMAL TRANSFER IN MHD FLOW TRAIN (MAIN COMBUSTOR THROUGH DIFFUSER)
- SPECIFIED DUTIES OR TEMPERATURE RANGES FOR HRSR COMPONENTS
- DATA FOR STEAM PLANT MODEL AS DESCRIBED IN REFERENCE 11
- MISCELLANEOUS COMPONENT EFFICIENCIES (E.G., INVERTERS, PUMPS, COMPRESSORS)
- LOSSES TO AMBIENT AS PERCENTAGE OF INPUT THERMAL ENERGY
- INITIAL ESTIMATE OF MASS AND ENERGY BALANCES*

OUTPUTS

- COAL PILE TO BUS BAR PLANT EFFICIENCY
- STEAM PLANT THERMODYNAMIC EFFICIENCY
- TEMPERATURE, PRESSURE ENTHALPY AND MASS FLOW RATE FOR EACH STREAM AS IT EXITS EACH NODE SHOWN ON THE SYSTEM DIAGRAM
- ENERGY TRANSFERS FROM STREAM TO STREAM AND TO AMBIENT AT EACH NODE

*MOST CONVENIENTLY THIS INITIAL ESTIMATE IS THE OUTPUT FROM A PREVIOUS CALCULATION

*Gross power less power required for oxidizer compression and O_2 production, if any.

The requisite Mollier charts for the combustion product state variables were calculated using the GE Coal Combustion Equilibrium Code (CCE), see Appendix C. The charts were stored in tabular form and used with a double interpolation subroutine. Some sample charts in graphical form are shown in Figures 1.6-1, 1.6-2, 1.6-3 and 1.6-4. Figures 1.6-1 and 1.6-2 are for seeded combustion products of Montana Rosebud (MR) coal dried to 5% moisture with no O_2 enrichment, before and after final oxidation. Figure 1.6-3 is also for MR coal but with 40% O_2 enrichment. The ratio of seed to coal was kept constant hence seed flow increases from 1% to 1.26% of combustion product mass flow rate. Finally, Figure 1.6-4 is for seeded combustion products of Illinois #6 coal dried to 2% moisture without O_2 addition.

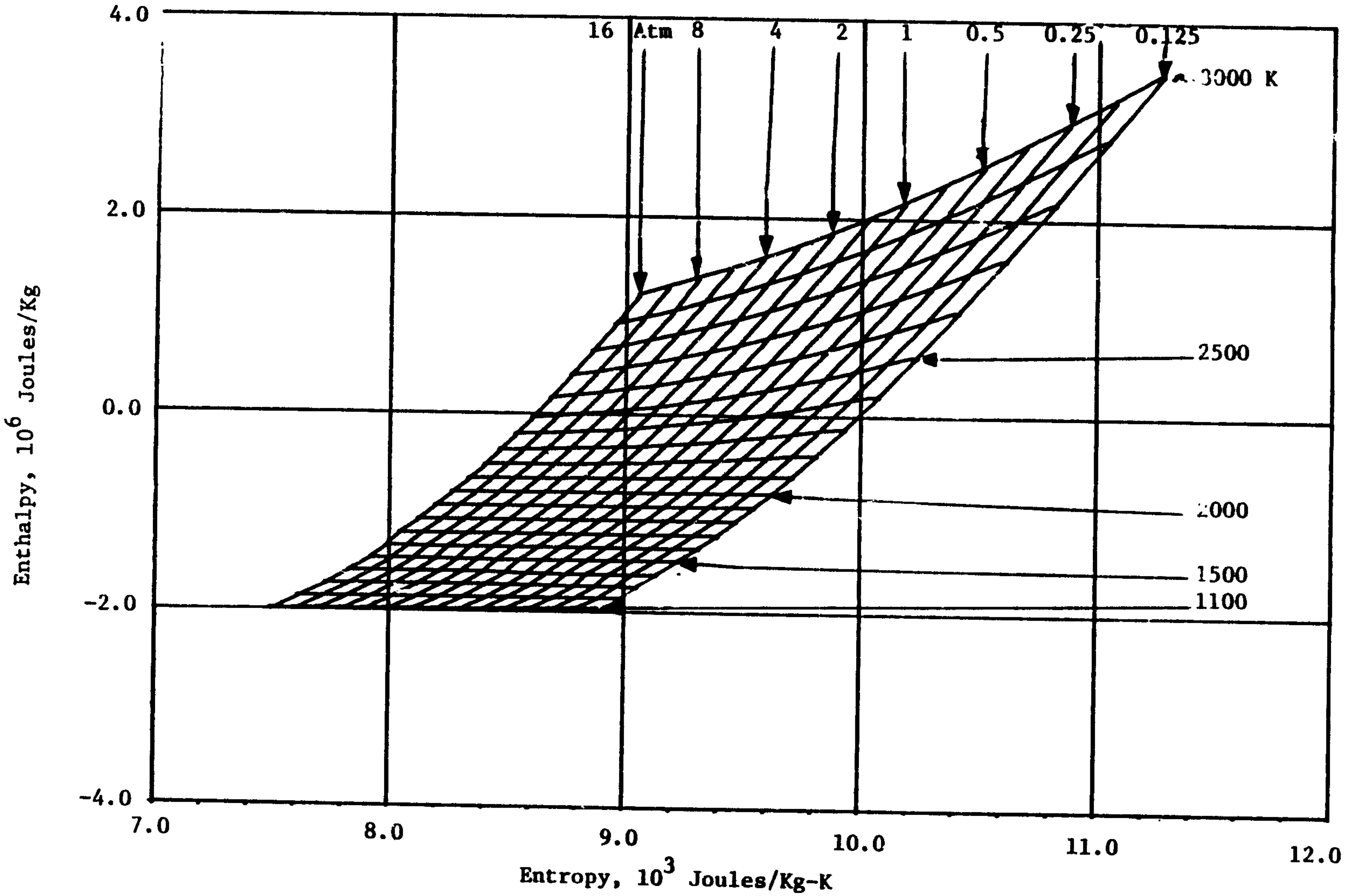


Figure 1.6-1. Mollier Chart Case 1.2, Montana Rosebud, 5% Moisture, 85% Slag Rejection, 90% of Stoichiometric Oxidizer, no O_2 Enrichment, Potassium in Seed 1% of Total Mass Flow. Applied to MHD Flow before Final Oxidation

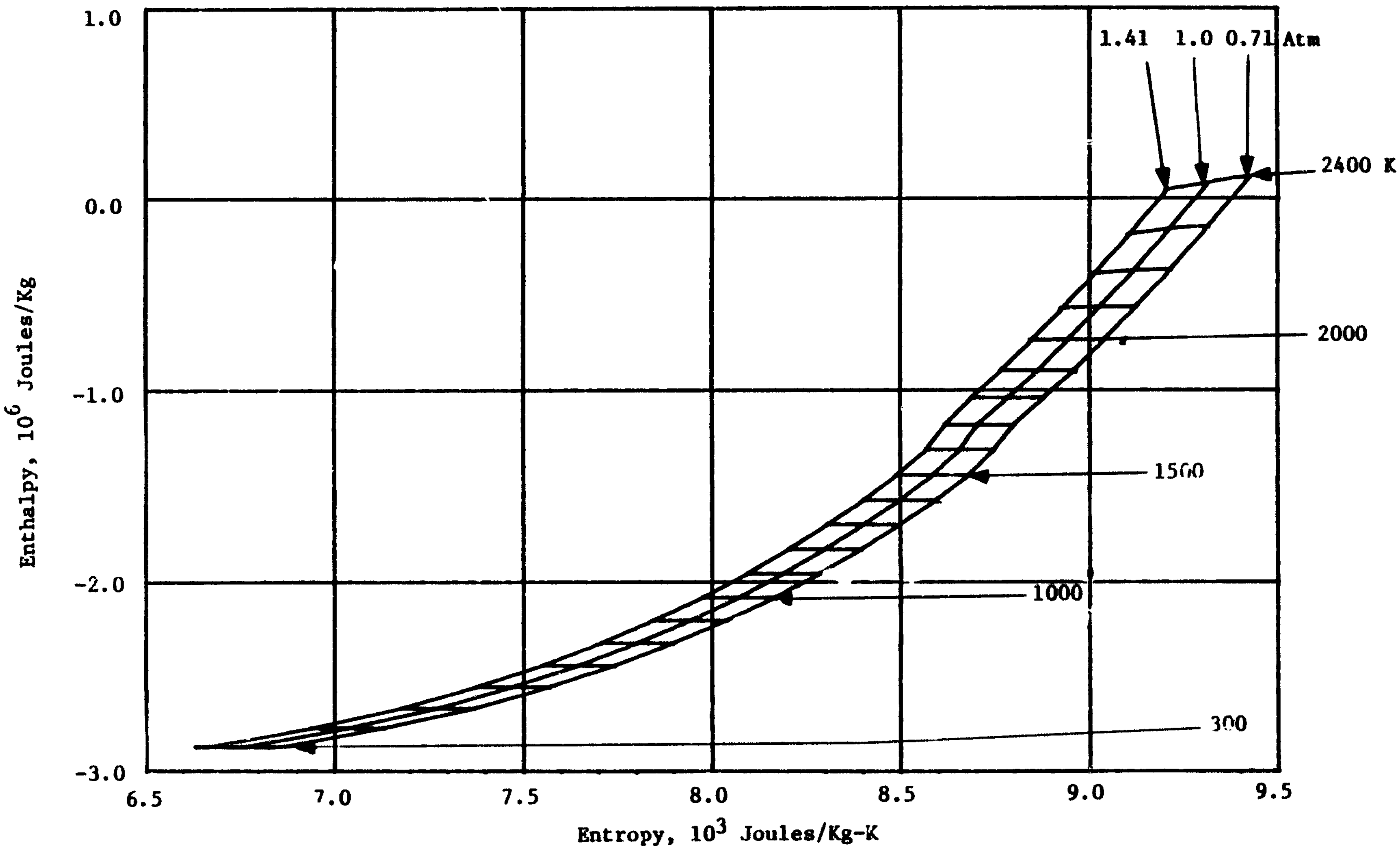


Figure 1.6-2. Mollier Chart, Case 1.2, Montana Rosebud, 5% Moisture, 85% Slag Rejection, 105% of Stoichiometric Oxidizer, No O_2 Enrichment. Applied to MHD Flow After Final Oxidation.

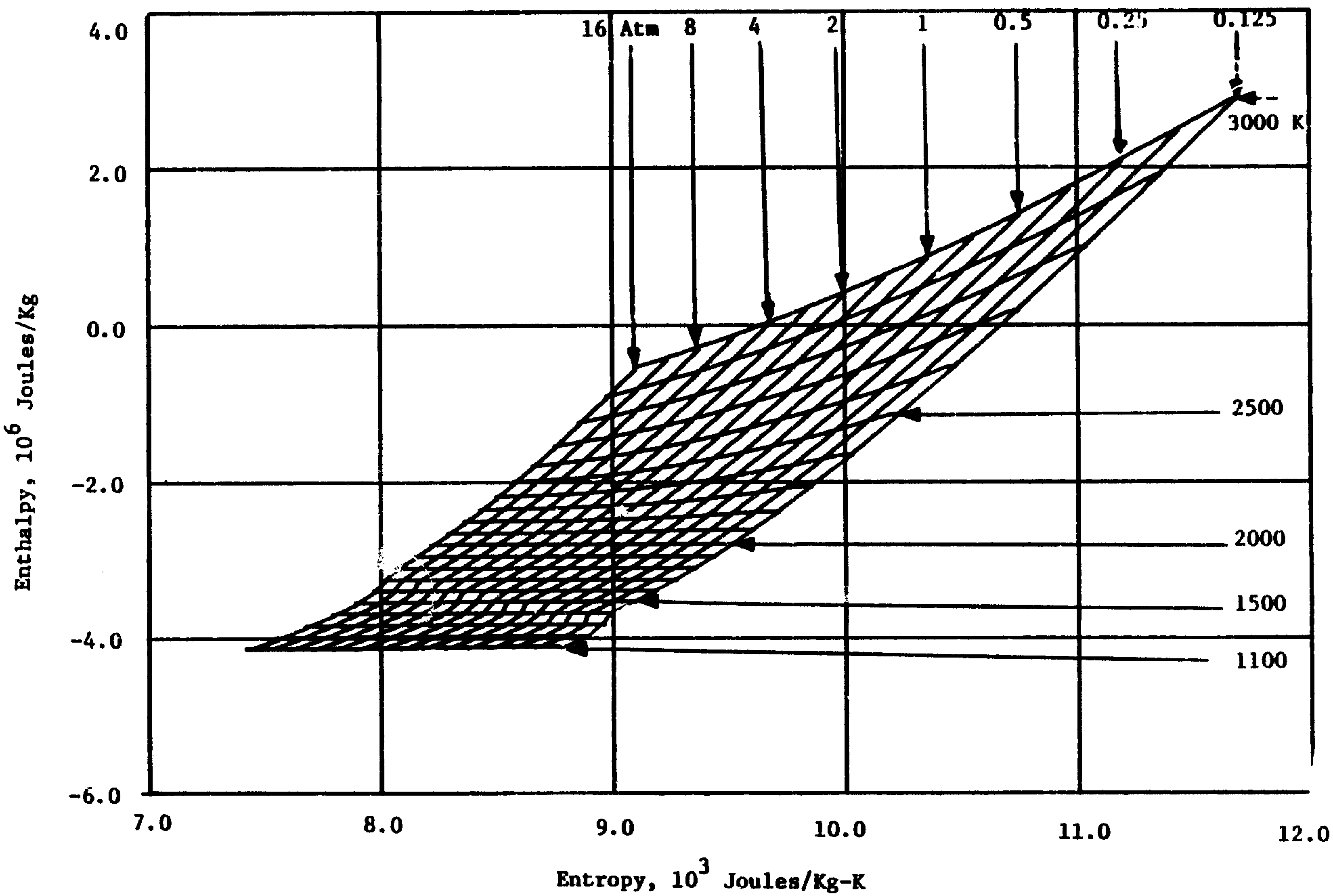


Figure 1.6-3. Mollier Chart Case 3.2, Montana Rosebud 5% Moisture, 85% Slag Rejection, 90% of Stoichiometric Oxidizer, 40 lb O_2 Added per 100 lb Air, Potassium in Seed 1.26% of Total Mass Flow. Applied to MHD Flow Before Final Oxidation.

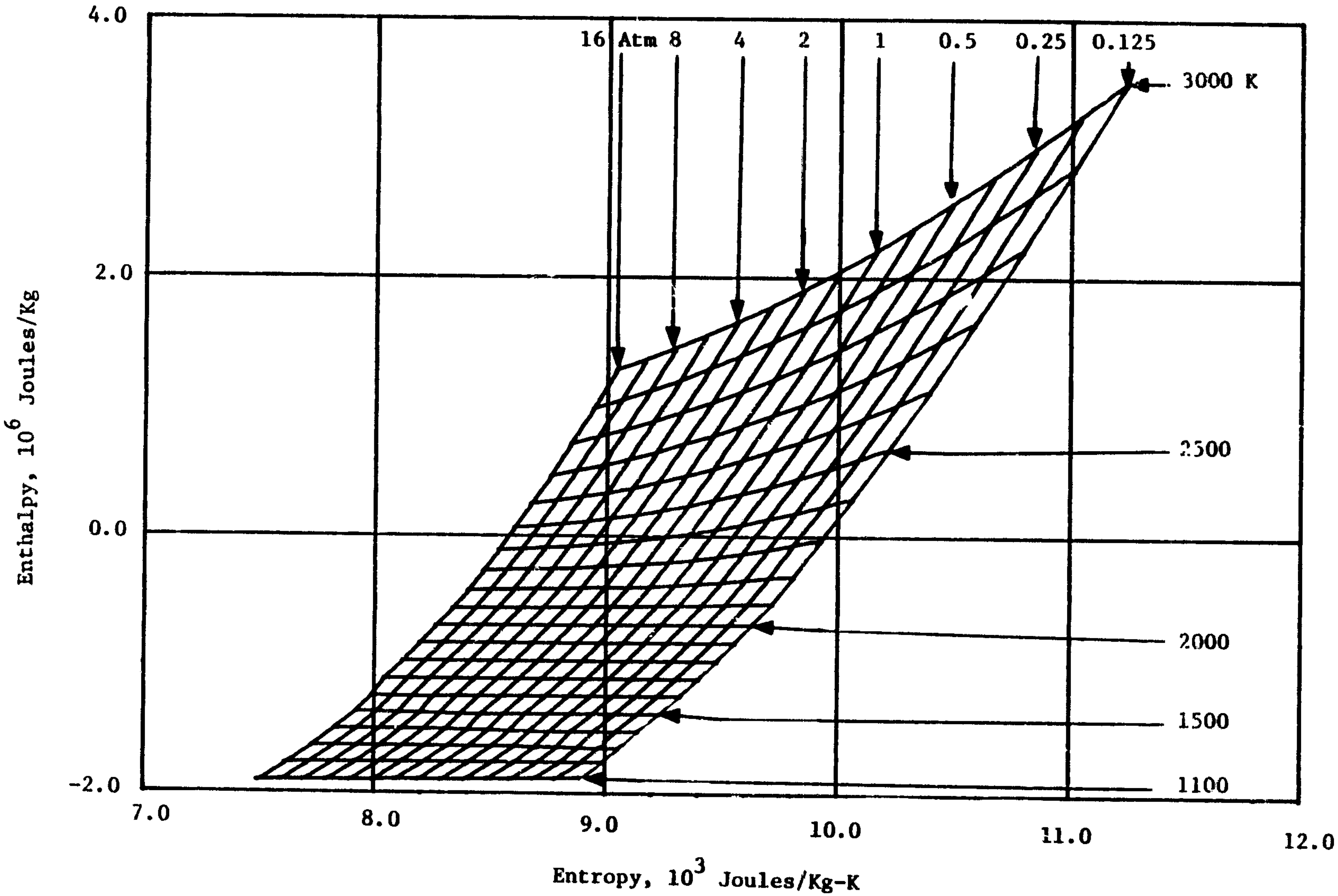


Figure 1.6-4. Mollier Chart Case 2.1, Illinois No. 6, 2% Moisture, 70% Slag Rejection, 90% of Stoichiometric Oxidizer, No O_2 Enrichment, Potassium in Seed 1% of Total Mass Flow. Applied to MHD Flow Before Final Oxidation

SECTION 2

SYSTEMS CONFIGURATIONS AND RESULTS

SECTION 2

SYSTEMS CONFIGURATION AND RESULTS

Configurations and results for the three base cases are presented in this section. Because it was anticipated that Base Case 2 (3000 F air preheat with no O_2 enrichment) would yield the best performance, most of the parametric variations were made about that case. Some of the summary graphs plotted for Base Case 2 also include data for Base Cases 1 and 3. For each reference case and for several alternative system configurations in Base Case 2, code OCSYS was used to generate a complete system balance. A corresponding system diagram was prepared along with a list of key state points.

The system diagrams show the mnemonic alphanumeric symbols referring to each component which were used in the system code. Most of the symbols are self-explanatory (e.g., MD = MHD generator) but a complete dictionary is included as Appendix A. Numbers on the system diagrams indicate locations at which state point data are tabulated.

Parametric variations from these basic system configurations were evaluated adjusting the energy flows. For all cases an energy flow summary was prepared. A complete set of system diagrams, state point tables and energy flow summaries are collected in Appendix B.

2.1 BASE CASE 1

2.1.1 REFERENCE SYSTEM DESCRIPTION

The plant arrangement and state points for the reference system Case 1.0 are given in Figure 2.1-1 and Table 2.1-1, respectively.

Referring to the MHD flow train in Figure 2.1-1, the combustor (CB1) is a two-stage cyclone with 85% slag rejection, exhausting into the MHD components, e.g., the nozzle (NZ), MHD channel (MD) and diffuser (DF). The remainder of the nodes in the gas path, from the diffuser exit to the stack, represent the HRSR subsystem. The radiant furnace is node RB, and the final oxidation furnace is the group of nodes SHH, FOF1 and FOF2, where SHH is the superheat panels in the top of the furnace, FOF1 the waterwall above the secondary air injection location and FOF2 the waterwall below. The convective pass is made up of nodes LPAH (air heater for the HTAH combustor), SHL (convective pass waterwall), RHH (high temperature reheat bundle) and RHL (low temperature reheat bundle). The back pass consists of the high and low economizers (ECH and ECL, respectively), the secondary air heater (SAH) and an electrostatic precipitator (ESP). The radiant furnace is assumed to be a balanced draft unit, (e.g., an inlet gas pressure of 14.7 psia), requiring an induced draft fan (IDF) to boost the exit gas pressure from the HRSR subsystem back to atmospheric.

In the present analysis, recirculation of flue gas from the exit of the ESP to the exit of the final oxidation furnace was not included (see Section 3.7). In the cases evaluated, the exit gas temperature from the final oxidation furnace was typically quite close to the desired range of 1800-1900 F. Neglecting gas recirculation to adjust the inlet gas temperature to the convective pass does not affect the system performance, but rather only the gas-side

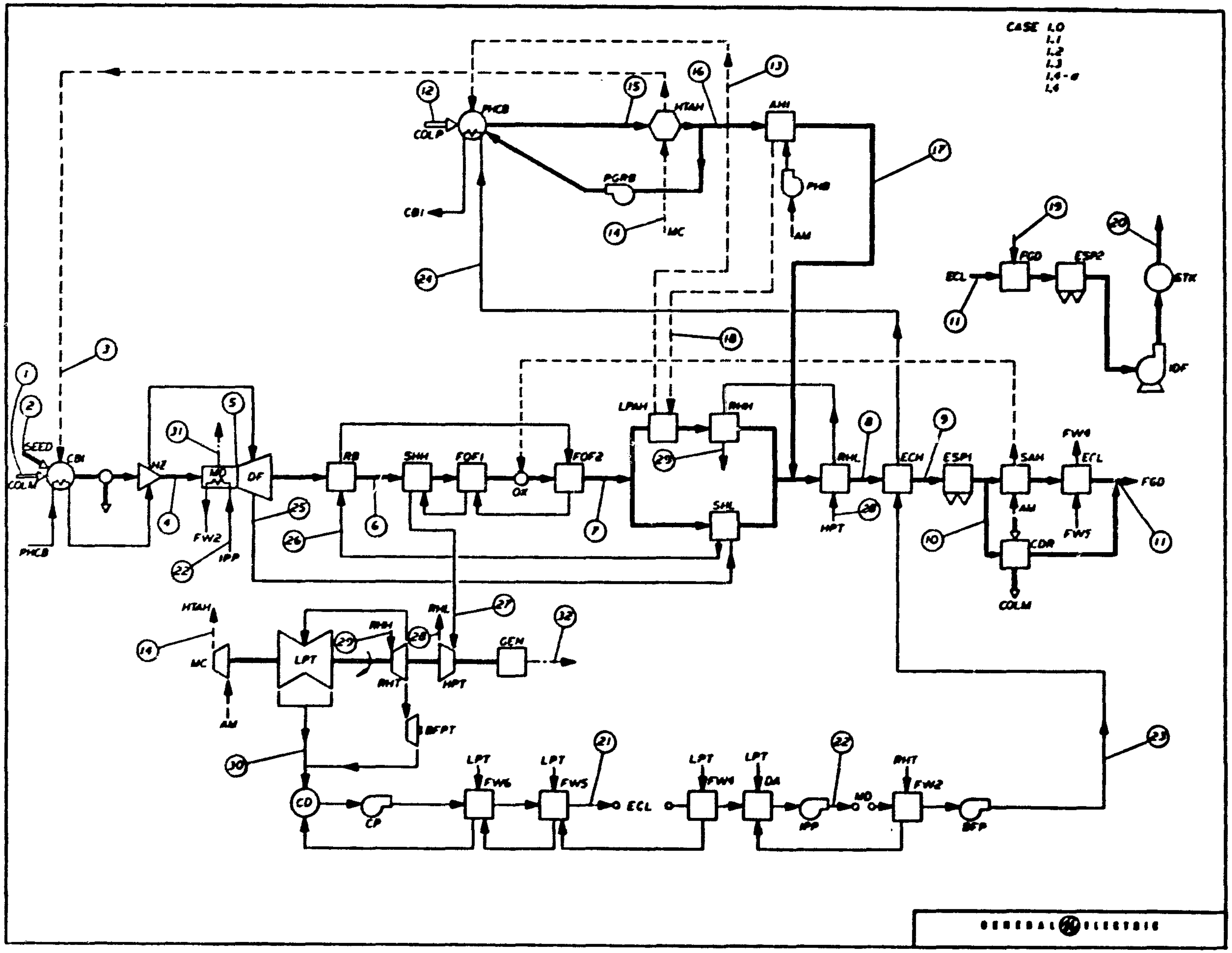


Figure 2.1-1. Plant Arrangement, Base Case 1

Table 2.1-1. State Points, Case 1.0

Location	T(F)/(K)	P(psia)	M(Kg/S)	E (MW)
1	220/378	-	79.40	2042.0
2	220/378	-	14.28	0.93
3	2700/1756	168.7	458.0	766.3
4	4800/2923	153.4	544.08	2638.8
5	3670/2295	17.60	544.08	1842.7
6	2900/1866	13.94	544.08	1474.5
7	1833/1274	13.82	643.65	874.5
8	762/679	13.66	964.66	599.9
9	513/540	13.56	964.66	449.2
10	513/540	13.56	233.65	109.7
11	336/442	13.43	967.47*	347.3
12	77/298	-	40.33	838.5
13	1100/867	20.0	189.9	116.2
14	688/638	174.9	458.0	165.5
15	3367/2126	17.97	371.66	927.7
16	1100/867	16.17	321.01	242.7
17	774/685	16.12	321.01	174.1
18	728/660	20.0	189.9	73.2
19	59/288	13.43	31.93	0.0
20	221/378	14.75	999.4	363.1
21	190/361	157	544.4	-
22	301/423	400	597.3	-
23	427/493	3900	597.3	-
24	527/548	3880	597.3	-
25	655/619	3670	597.3	-
26	715/653	3610	597.3	-
27	1000/811	3500	597.3	-
28	604/591	768	586.9	-
29	1000/811	691	591.8	-
30	106/314	2.3"Hg	448.2	-
31	-	-	-	654.2
32	-	-	-	644.3

*Includes Coal Drying Moisture

temperature distribution and thus the required heat transfer area (provided that the gas is recirculated from above the pinch-point).

The high temperature air heater (HTAH) system for Case 1.0 is described in Section 3.2. Briefly, it uses an atmospheric pressure combustion system with a Wellman-Galusha gasifier (not shown) and a combustor located in the heat exchanger dome (node PHCB). Air is delivered by blower PHB to the preheat combustor via air heaters AH1 and LPAH in the preheat and main combustion gas streams, respectively. Air to the HTAH is delivered directly from the main compressor (MC). Downstream of AH1, the HTAH combustion gas is mixed with the main stream. The mixed stream delivers energy to reheater RHL and economizers

ECH and ECL as well as secondary air heater SAH. Sulfur removal is by dry scrubbing (node FGD), see Section 3.8.

Heat for the coal dryers (node CDR) is supplied by a combustion gas stream extracted from the HRSR subsystem at the exit of the ESP.

In the steam cycle, water from the condenser is first raised to 190 F in two feedwater heaters, passed through the low temperature economizer (ECL), and into the deaerator, exiting at 300 F. At this point the water pressure is raised to 400 psi and used to cool the MHD channel, typically raising the water temperature 75-100 F. A final stage of feedwater heating results in an outlet water temperature from the feedwater heat train of 427 F. The water then passes through the water/gas pinch-point at the gas outlet of the high temperature economizer. From the high temperature economizer, the water cools the gasifier in the HTAH system, and then the remaining MHD components (combustor, nozzle, diffuser). As detailed in Section 3.7, the water is subsequently raised to the high pressure turbine inlet conditions in the HRSR components and then reheated for the reheat turbine.

2.1.2 PARAMETRIC VARIATIONS

Parametric variations are summarized in Table 2.1-2. Without changing the plan configuration, Case 1.1 evaluates the effect of using Illinois #6 coal. In Case 1.2, the oxygen enrichment of the main combustion air was removed in order to determine the effect of using air only. For Case 1.3, the maximum magnetic field was near the generator entrance, tapering to 6T near the generator exit (for Case 1.0, 6T tapered to 5T). The increase in magnetic field was limited to 7T (as opposed to 8T in Base Case 2) to avoid going to niobium-tin superconductor, consistent with the near state-of-the-art nature of Base Case 1. Case 1.4 introduced a single-stage vortex combustor with 85% slag rejection, replacing the cyclone. This was also compared with Case 1.4a in which only 70% slag rejection is assumed in the single-stage vortex.

Table 2.1-2. Base Case 1 Single Parameter Variations
(Reference Case Efficiency, 41.4%)

Reference	Parametric Case	Variation	ΔEFF
MR Coal	1.1	I6 Coal	+1.49
Air + 10% O_2	1.2	Air Only	+0.32
(6-5) TESLA	1.3	(7-6) TELSA	+0.80
2-Stage Cyclone Combustor, 85% Slag Rejection	1.4	Single Stage Vortex, 85%	+0.35
	1.4(a)	Single Stage Vortex, 70%	+0.09

2.1.3 RESULTS AND DISCUSSION

The effects of the parametric variations on overall plant efficiencies for Base Case 1 are shown in Table 2.1-2. There is no penalty for seed reprocessing since sulfur capture is by dry scrubbing (Section 3.8) for which energy requirements were included in the system heat balance. The reference HTAH subsystem (see Section 3.2) is near state-of-the-art, limited to 2700 F preheat. As a consequence of this technology constraint and the higher heat loss associated with the atmospheric pressure HTAH gasifier, the performance of all systems in Base Case 1 is low. Overall plant efficiencies range from 41.41 - 42.90%.

2.1.3.1 Illinois #6 Coal vs Montana Rosebud

Changing coals had the strongest effect on plant efficiency of all the variations considered in Base Case 1. The dryer Illinois coal resulted in improved channel performance and lower coal drying penalty, i.e., thermal power losses. As a result, the plant efficiency increased 1.49 points.

2.1.3.2 Air Only vs Oxygen Enrichment

Deleting the oxygen enrichment resulted in two compensating effects. The removal of the oxygen lowered the generator performance. However, that left additional thermal energy to be converted in the steam plant. Coupled with the reduced plant internal power without the oxygen plant load, there resulted a very slight net increase in the efficiency of 0.32 points.

2.1.3.3 Magnetic Field

The result of increasing the magnetic field was an increase in the channel enthalpy extraction from 24.79% to 26.53%, which in turn boosted the plant efficiency by 0.80 points. For a more detailed description of the effect of channel performance on the plant efficiency, see Section 2.2.3.

2.1.3.4 Single-Stage Vortex vs. Two-Stage Cyclone

Overall plant efficiency is affected by two combustor parameters, heat loss and slag rejection. Holding slag rejection constant, the single-stage vortex case cuts the combustor heat loss and plant efficiency rises 0.35 points. However, decreasing the slag rejection in the single stage combustor to 70% has an adverse effect on the channel performance and plant efficiency is then essentially identical to the reference case (two-stage cyclone with 85% slag rejection).

2.2 BASE CASE 2

2.2.1 REFERENCE SYSTEM DESCRIPTION

The plant arrangement and state points for the reference system, Case 2.0, are given in Figure 2.2-1 and Table 2.2-1, respectively. The reference case (2.0) component arrangement was selected to represent a simple, or straight-forward, system with an indirectly-fired HTAH. As such, it is not a high performance configuration, but rather serves as a baseline against which the many parametric variations can be compared and the effect of these variations clearly evaluated.

With the exception of the combustor, the general arrangement of the MHD flow train is identical to that for Case 1.0, Section 2.1.1, upstream of the convective passes. Case 2.0 uses a single-stage vortex with 85% slag rejection. The metallic air heater in the HRSR is absent in the reference case but has been incorporated in some of the alternative configurations. The preheat combustor combustion product stream is separately treated for sulfur removal by dry scrubbing since for the MHD combustion products sulfur is captured by the potassium seed.

The high temperature air heater (HTAH) system for Case 2.0 is described in Section 3.2. Briefly, it uses a pressurized combustion system with a two-stage cyclone combustor (nodes PHCB and CB2). Air is delivered, both to the HTAH and to the preheat combustor, directly from their respective compressor (MC and PHC) outlets. Downstream of the HTAH, the pressure of the combustion gas is let down through a gas turbine which drives the air compressor for the preheat combustor.

Other parts of the system are also similar to Case 1.0 with the following exceptions. An additional coal dryer CDR2 has been included to utilize the low grade thermal energy remaining in the preheat combustor flue gas at the gas turbine exit. Two final stages of feedwater heating result in an outlet water temperature from the feedwater heater train of 510 F. The water then passes through the water/gas pinch-point at the gas outlet of the high temperature economizer. From there, the water cools a second small economizer and the first stage combustor in the HTAH system, and then the remaining MHD components.

2.2.2 PARAMETRIC VARIATIONS*

Case 2.0(s) evaluates the effect on channel performance and system efficiency of reducing slag rejection in the MHD combustor from 85 to 70 percent.

Cases 2.0(a) and 2.0(b) examine system configurations which use gasifiers to supply fuel-gas to both the MHD and preheat combustors. In both cases, a split-stream gasifier (Sections 3.1.2.3 and 3.1.2.4) is used to supply a hydrogen-rich gas to the preheat combustor and a carbon-rich gas to the MHD combustor. The two fuel-gas streams from the split-stream gasifier have approximately equal energy flow rates and since the MHD

*Systems diagrams and state point tables for parametric cases are collected in Appendix B.

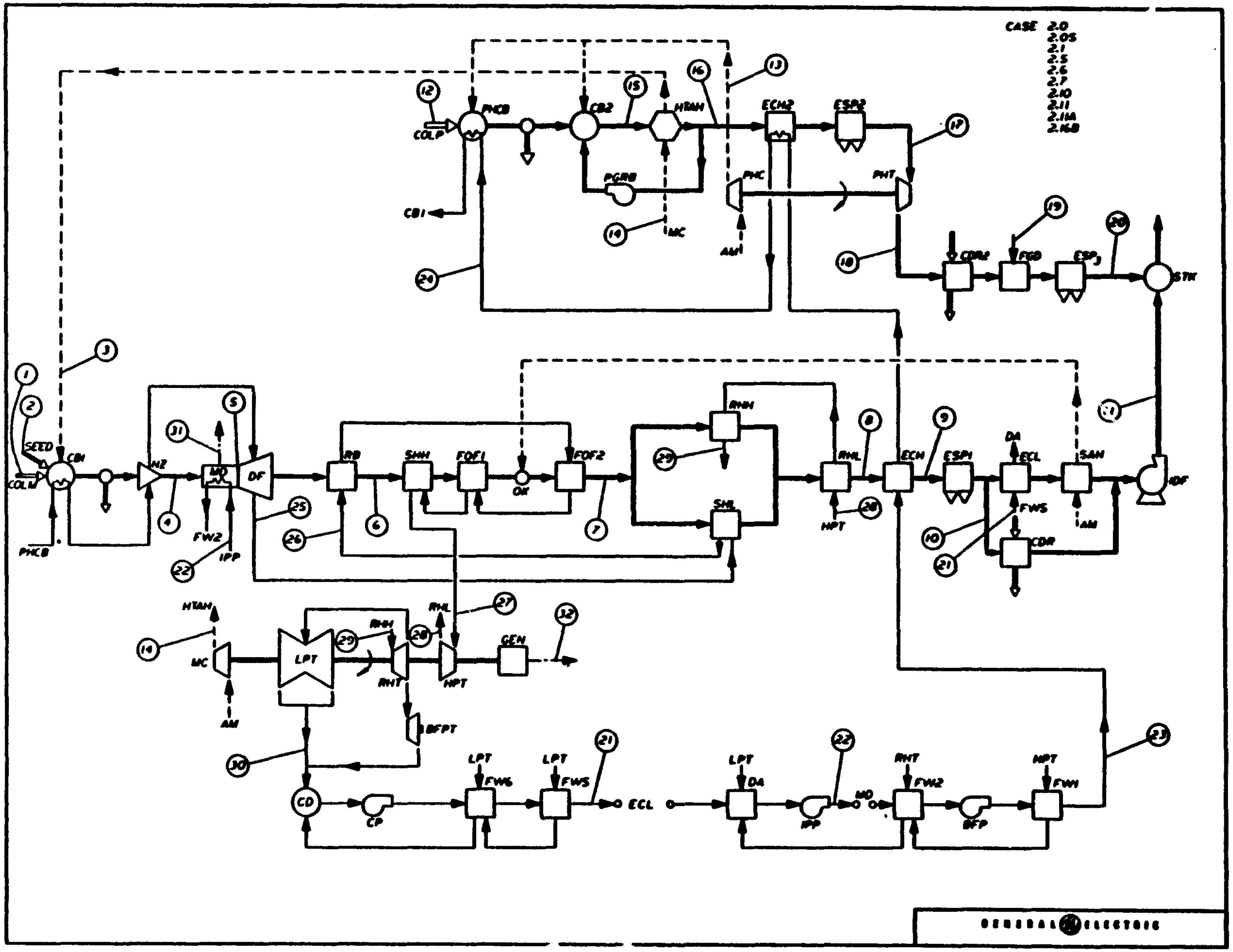


Figure 2.2-1. Plant Arrangement, Case 2.0 and Some Parametric Cases

Table 2.2-1. State Points, Case 2.0

Location	T (F)/(K)	P (Psia)	$\dot{m}$ (Kg/s)	$\dot{E}$ (MW)
1	220/378	-	70.36	1809.2
2	220/378	-	12.7	0.8
3	3000/1922	136.5	529.4	995.6
4	4677/2854	127.0	605.7	2668.4
5	3582/2246	17.6	605.7	1881.3
6	2900/1867	14.55	605.7	1533.7
7	1867/1293	14.44	693.9	875.9
8	952/784	14.27	693.9	439.1
9	611/595	14.17	693.9	288.4
10	611/595	14.17	250.1	104.7
11	260/400	14.75	678.0*	137.1
12	220/378	-	42.19	1084.8
13	648/616	148.5	359.7	121.4
14	623/602	141.9	529.4	170.7
15	3300/2089	136.5	487.8	1167.7
16	985/803	131.1	398.1	269.9
17	877/743	130.7	397.4	241.6
18	373/463	15.15	397.4	117.7
19	59/288	15.15	13.1	0.0
20	162/346	14.75	410.5	101.7
21	190/361	157	564.7	-
22	301/423	400	672.8	-
23	510/534	4230	672.8	-
24	608/593	4200	672.8	-
25	694/641	3950	672.8	-
26	720/656	3850	672.8	-
27	1000/811	3500	672.8	-
28	603/591	768	662.2	-
29	1000/811	691	662.2	-
30	106/314	2.3" Hg	462.7	-
31	-	-	-	658.3
32	-	-	-	661.7

*Excludes coal drying moisture

combustor requires a larger thermal input than the preheat combustor, augmentation of the MHD combustor input is required. In Case 2.0(a), the thermal input to the MHD combustor is supplemented with a coal-fired single stage vortex combustor. In Case 2.0(b), the additional thermal input to the MHD combustor is supplied by a single-stage gasifier. Note, in the system diagrams, that the gasifiers are cooled with steam, immediately after it has passed through the transition region in the convective pass waterwall. In a supercritical steam cycle, the steam at this condition still has very high density, sufficient for cooling the gasifier beds.

Case 2.1 evaluates the effect of using Illinois #6 coal in the reference plant configuration. A single stage combustor with 70% slag rejection in the MHD combustor was assumed.

Case 2.2(a) employs a two-stage cyclone with 85% slag rejection for the MHD combustor. The difference here, as compared to 2.0, is in the combustor heat loss. Case 2.2, in addition to a 2-stage cyclone main combustor, has a "hot-bottom" HTAH (which is discussed below under Case 2.16).

Case 2.4 was set up as a method of comparing the NASA LeRC and GE MHD generator analyses and channel optimization procedures. In this case, NASA LeRC supplied GE with an optimized set of channel parameters, calculated under an agreed upon set of conditions. These conditions were a 2-stage cyclone MHD combustor (with 85% slag rejection and the GE-specified heat loss) and a 20 m channel length. The corresponding GE calculation is designated Case 2.4(a).

Case 2.5, 2.6 and 2.7 consider variations in components which influence channel performance for the reference plant configuration. In Case 2.5, Cs rather than K was used as the seed element for the MHD working fluid. Although the use of Cs will probably modify seed behavior and thus influence design of the HRSR subsystem, these considerations were not addressed in the present study. In Case 2.6, a supersonic rather than subsonic channel design was used. In Case 2.7, the magnetic field was assumed to have a value of 8T at the generator entrance, and to taper to 7T at the generator exit (as compared to 6T tapered to 5T in Case 2.0).

Cases 2.10, 2.11 and 2.11(a) evaluated the effect of size, e.g., thermal input to the MHD combustor, on generator performance and overall plant performance. In these cases, the only component whose characteristics were assumed to be size dependent was the MHD channel. All other components, i.e., combustors, compressors, steam plant, etc., were assumed to have the same efficiencies as in the reference case. Case 2.10 is a plant with a 1500 MWt input to the combustor (53.5% of the reference value of 2800 MWt) and Case 2.11 is for a 2000 MWt input (71.4% of the reference value). Case 2.11a, in addition to reducing the thermal input to 2000 MWt, assumed 70% slag rejection. That is, this case evaluated the effect of slag rejection in a smaller (than the reference) size.

Case 2.12 replaces the pressurized HTAH combustion system with an atmospheric pressure system. In addition, the two-stage steam-cooled cyclone preheat combustor was replaced, because it has unacceptably high heat losses at atmospheric pressure (due to its large size and resultant surface area). The alternate combustor selected was a single stage,

regeneratively air-cooled, cyclone (presently under development by General Electric). The preheat combustor air is first heated to 600 F in a metallic air heater downstream of the HTAH, and then delivered to the preheat combustor cooling channels, where its temperature is elevated to 1200 F before admission to the combustion chamber.

Because of the atmospheric pressure of the reheat flow, a considerably larger heat transfer area is required in the HTAH's compared to pressurized operation. This condition implies an increased surface area for the HTAH vessels, and potentially a larger heat loss from the HTAH system. However, it was assumed that the atmospheric HTAHs were more highly insulated than the pressurized units, and the heat loss was held constant for the two designs (5% of the input chemical and sensible heat to the dome-mounted combustors).

Case 2.16 is a high performance configuration, that includes a "hot bottom" HTAH and a highly optimized channel. The MHD combustor air preheat system is one in which the air temperature is first raised to 1300 F in a metallic heat exchanger in the HRSR subsystem, and is then raised to 3000 F in the HTAH. This arrangement regenerates a significant amount of heat from the MHD flow train, thereby reducing the fuel requirement for the preheat combustor. This fuel requirement is further reduced by regenerating heat with the preheat combustor air from the combustion gas down-stream of the HTAH. The MHD channel is one in which the magnetic field intensity is varied along the length of the channel to produce a constant transverse electric field, $E_y = 4$ kVm. Case 2.16(a) evaluates the effect of the high performance air preheat system alone, and Case 2.16(b) the $E_y = 4$ kV/m channel alone.

Case 2.17 uses a chemically active pressurized fluidized bed (CAPFB) gasifier as the first stage of the preheat combustor, thereby examining an alternative to flue gas desulfurization for sulfur control in the preheat combustor flow train. The gasifier also eliminates the need for an ESP before the gas turbine, as the gasifier is presumed to have very low ash carry-over. The off-gas from the gasifier is used to raise the second stage preheat combustor air temperature to 818 F before admission to the second stage. The gasifier air supply uses an intercooled compressor, as the gasifier design (by Foster Wheeler) assumed a 404 F air inlet temperature. As was the case with the SPMB and S^3PMB gasifiers (Cases 2.0(a) and 2.0(b)), the CAPFB is cooled with (low superheat) steam.

Case 2.18 considers an alternate method to the hot bottom HTAH for regenerating heat from the MHD flow train. Here, the air and recirculated flue gas for the preheat combustor are heated to 1200 F and 1300 F, respectively (in metallic heat exchangers), in the HRSR subsystem before admission to the combustor*. The preheat combustor is a pressurized version of the combustor used in Case 2.12, e.g., single-stage, regeneratively air-cooled. In this pressurized configuration, the heat loss from the combustion chamber to the air coolant is significantly reduced (as compared to the atmospheric combustor), causing only a 100 F rise in air temperature in the cooling channels, versus 600 F for atmospheric operation. The decision to use this single stage combustor does not imply unacceptability of the (reference)

*The air is further heated to 1300 F in cooling the combustor before admission to the combustion chamber.

two-stage cyclone, but rather a desire to evaluate performance of the pressurized, single-stage, regeneratively air-cooled combustor somewhere within Base Case 2.

2.2.3 RESULTS AND DISCUSSION

The overall plant efficiencies (both with and without seed reprocessing) for the 21 cases in Base Case 2 are shown in Table 2.2-2. Included in Table 2.2-2 is a "best" case, which is a plant configuration that incorporates the parametric variations (from the reference case) that have a positive influence on plant performance. The highest performance generator (Case 2.7) was not used because the electrical stresses in the channel were judged to be too high. The single stage, regeneratively air-cooled, combustor is recommended because of simplicity (compared to the two-stage cyclone), low cost and low heat loss. A system analysis for this configuration was not performed, but a preliminary estimate of performance was made based on the results of the single parameter variations discussed below.

Of the large number of configurations examined, there is a comparatively small variation in efficiency, from ~42.0% to - 45.6%, or a difference of 3.6 percentage points from worst to best. This small difference is primarily a result of the fact that all plant configurations were designed to have a high thermal efficiency in the sense that thermal losses are a small, approximately constant, fraction of the total coal energy input. The balance of the energy is input to the power cycle. The efficiency variations are therefore determined by the split in the energy input to the power cycle between the MHD channel and the steam cycle; the larger the fraction of the input energy converted in the channel, the higher the plant efficiency.

A simplified picture of the energy flow in an indirectly-fired HTAH plant is shown in Figure 2.2-2. The thermal losses from the plant is the sum

$$Q_L = (1 - \beta_1)(1 - \eta_p)(Q_{c1} + \alpha Q_{c2}) + (1 - \alpha - \beta_2) Q_{c2}.$$

The remainder of the total input, $Q_c - Q_L$, is input to the power cycle

$$Q_{pc} = [\eta_p + (1 - \eta_p)\beta_1] Q_{c1} + \{\alpha [\eta_p + (1 - \eta_p)\beta_1] + \beta_2\} Q_{c2}$$

and is then split between the channel output

$$P_{MD} = \eta_p (Q_{c1} + \alpha Q_{c2}),$$

and the input to the steam cycle

$$Q_s = \beta_1 (1 - \eta_p) Q_{c1} + [\alpha\beta_1 (1 - \eta_p) + \beta_2] Q_{c2}.$$

In the Base Case 2 plants, the ratio Q_L/Q_c is approximately constant. The system efficiency is therefore determined by P_{MD}/Q_{pc}, increasing as this ratio increases.

Table 2.2-2. Overall Plant Efficiency*
Base Case 2

Case No.	Parameter Variation	Efficiency, % Without Seed Reprocessing	Efficiency, % With Seed Reprocessing
2.0	Reference System	43.66	43.45
2.0s	70% Slag Rejection	43.12	42.91
2.0a	S^3PMB + Coal 91% Slag Rejection	42.66	42.53
2.0b	S^3PMB + SPMB	43.42	43.22
2.1	Illinois #6 Coal 70% Slag Rejection	44.48	43.30
2.2	2-Stage Cyclone MHD Combustor Hot Bottom HTAH	44.13	43.89
2.2a	2-Stage Cyclone MHD Combustor	43.23	43.02
2.4	2-Stage Cyclone MHD Combustor NASA Generator 20 m Channel	43.47	43.26
2.4a	2-Stage Cyclone MHD Combustor E_y = 4 kV/m Generator 20 m Channel	43.45	43.23
2.5	Cs Seed	44.72	44.57
2.6	Supersonic Generator	43.01	42.81
2.7	8-7T Magnetic Field	44.91	44.69
2.10	1500 MWt MHD Combustor Input	41.98	41.78
2.11	2000 MWt MHD Combustor Input	42.91	42.70
2.11a	2000 MWt MHD Combustor Input 70% Slag Rejection	42.34	42.13
2.12	1 atm. HTAH Reheat Single Stage Preheat Combustor	43.00	42.79
2.16	Hot Bottom HTAH E_y = 4 kV/m Generator	45.49	45.25
2.16a	Hot Bottom HTAH	44.56	44.33
2.16b	E_y = 4 kV/m Channel	44.56	44.34
2.17	CAPFB Preheat Combustor	42.53	42.33
2.18	1200 F Air and 1300 F Recirc. Gas for Single Stage Preheat Combustor	44.26	44.03
Best	Illinois #6 Coal 85% Slag Rejection E_y = 4 kVm Generator Hot Bottom HTAH Single Stage Preheat Combustor	47.0 (EST)	46.3

*Efficiencies are quoted to the nearest 0.01% to minimize the effect of round off error on small differences but are not intended to indicate that level of absolute precision.

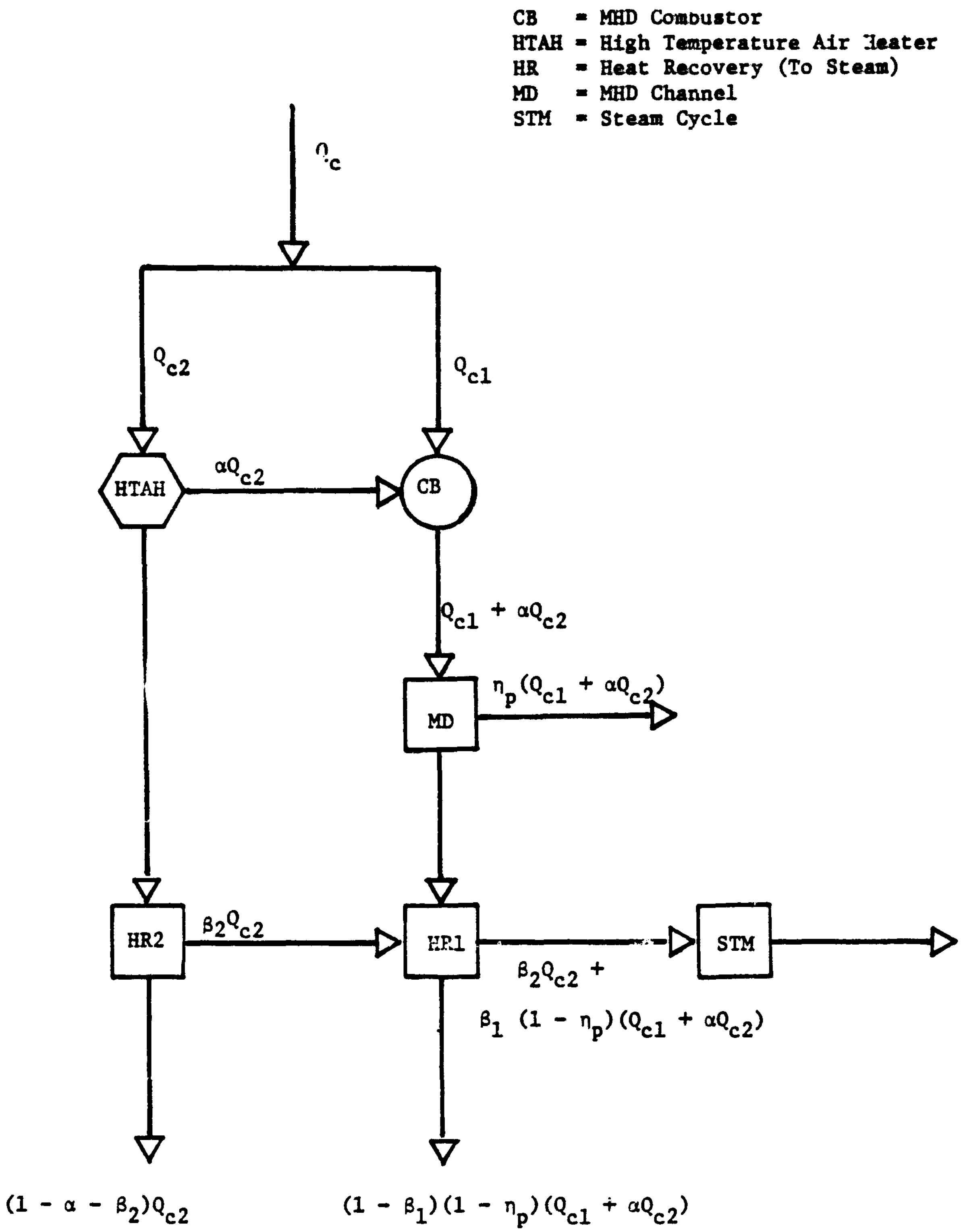


Figure 2.2-2. Energy Flow for a Plant with an Indirectly-Fired HTAH

This overview of the plant energy is quantified by the data shown in Figures 2.2-3 and 2.2-4. The ratio of the thermal losses divided by the total coal input is plotted versus the plant efficiency in Figure 2.2-3. Here, the thermal losses are defined as the sum of: stack loss, rejected solids, coal drying and other losses to ambient. The fractional losses for most cases are tightly grouped around 15%, with a few exceptions. The configurations with gasifiers (Cases 2.0(a), 2.0(b) and 2.17) and Case 2.12, the atmospheric HTAH combustion system, have higher losses. Case 2.4 (the NASA-specified generator) has a slightly lower than average loss, and Case 2.1 (Illinois #6 coal) is signficantly lower because of the reduced coal drying requirement.

The ratio of the gross MHD generator output divided by the total coal input is plotted versus efficiency in Figure 2.2-4. Here, most of the cases fall within a very tight linear band. Those few cases that scatter outside the band are the configurations that have exceptional heat losses (compare Figures 2.2-3 and 2.2-4). Figure 2.2-4 clearly illustrates the conclusion stated above: with losses that are a constant fraction of the coal input, the variation in efficiency is proportional to the fraction of the total input remaining that is converted to output in the channel. A 1-percentage point increase in (the normalized) channel output results in a 0.44-percentage point increase in plant efficiency.

With regards to overall plant efficiency, it is increased by the following (perhaps obvious) plant characteristics:

1. An increase in the fraction of the total coal input which enters the channel,
2. An increase in channel fractional enthalpy extraction,
3. A decrease in fractional plant losses.

By comparing selected cases to Case 2.0, the reference case, the effect on plant efficiency of varying a single parameter can be evaluated. These results are presented in Tables 2.2-3, 2.2-4 and 2.2-5.

2.2.3.1 Comparison of Two-Stage Cyclone and Single Stage Vortex Combustors

The single-stage and two-stage cyclone combustors are characterized by two parameters, heat loss and slag rejection, with 85% slag rejection and a single-stage heat loss taken as the reference performance parameters. Decreasing slag rejection in the single-stage combustor to 70% decreases efficiency by 0.54* percentage points in the 2800 MWt combustor, and 0.57 points in the 2000 MWt combustor. This decrease in efficiency is a result of decreased plasma conductivity due to the presence of the increased slag. Holding the slag rejection constant at 85% and increasing the combustor heat loss to that typical of a two-stage cyclone, the efficiency is decreased 0.43 points. This effect is also a result of decreased plasma conductivity, e.g., the increased heat loss lowers the inlet plasma temperature to the channel. Thus a single stage vortex combustor with 70% slag rejection and a

*See footnote Table 2.2-2.

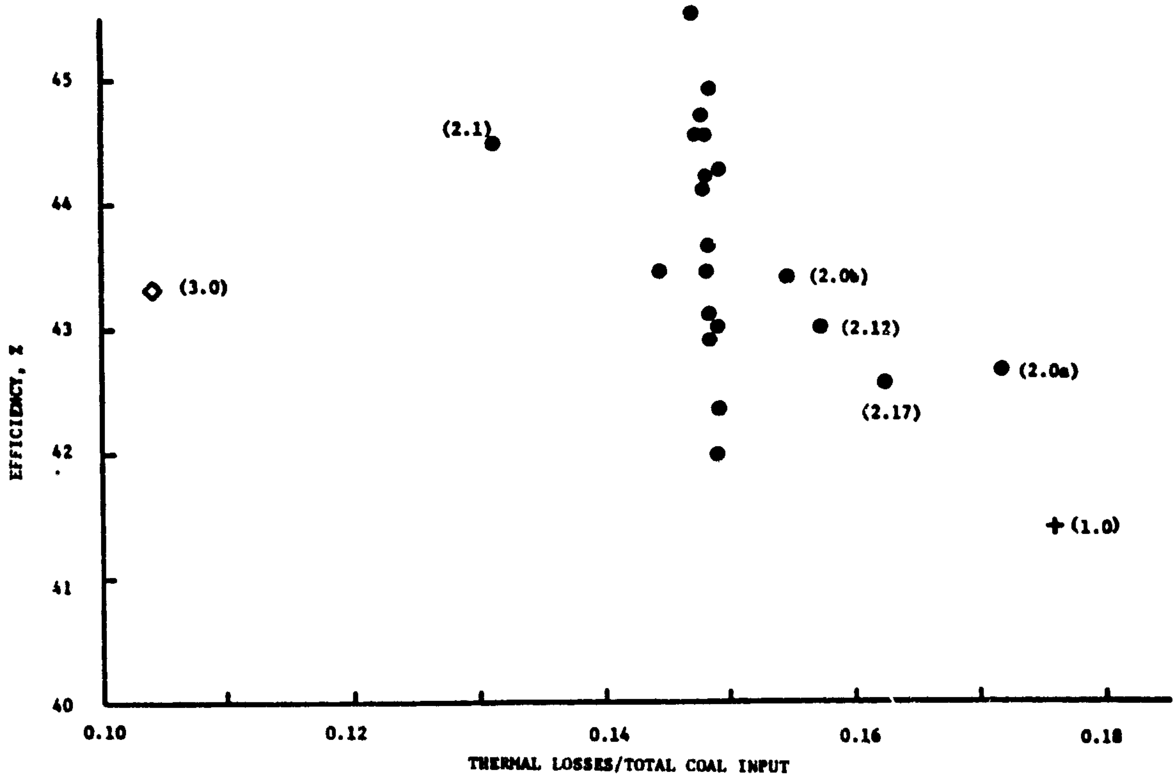


Figure 2.2-3. Plant Thermal Losses

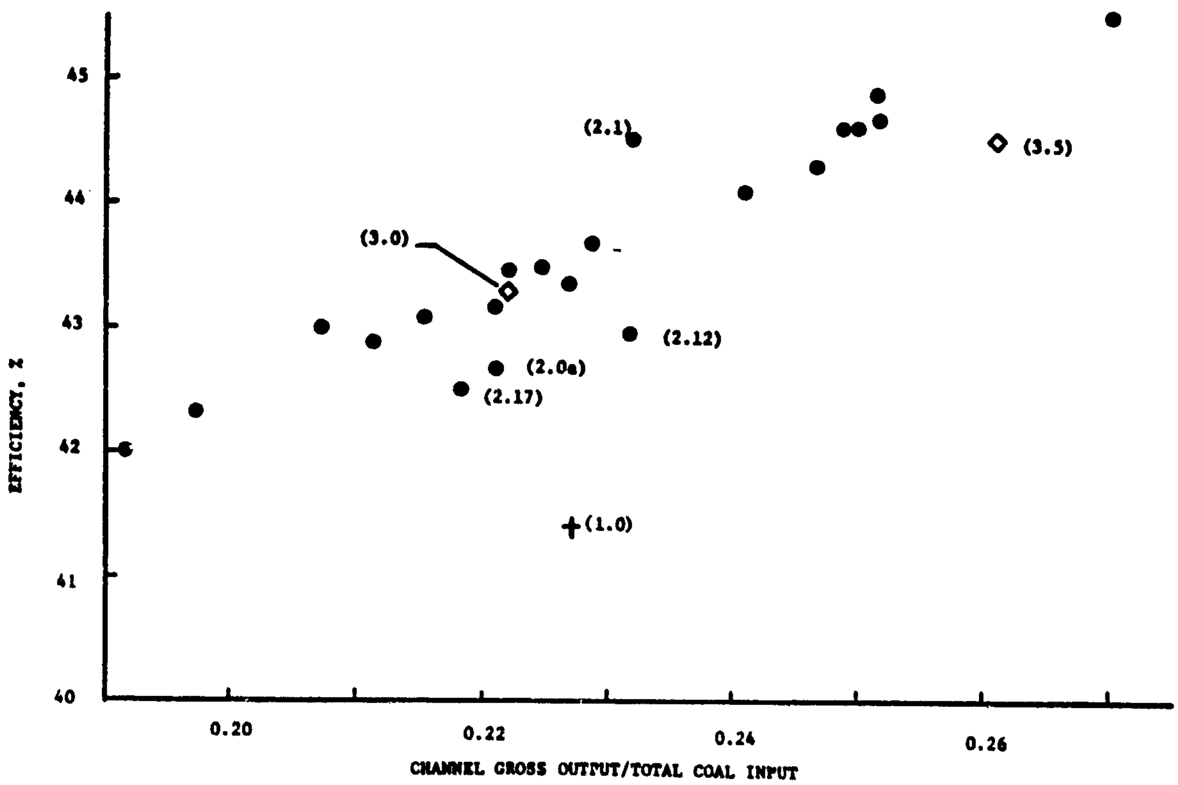


Figure 2.2-4. Relationship of Plant Efficiency to Channel Output

Table 2.2-3. Base Case 2 Single Parameter Variations MHD Combustor
(Reference Case Efficiency 43.5%)

REFERENCE	PARAMETRIC CASE	VARIATIONS	ΔEFF
SINGLE STAGE VORTEX WITH 85% SLAG REJECTION	2.0(s)	70% SLAG REJECTION	-0.54
	2.2(a)	2-STAGE CYCLONE	-0.43
	2.0(a)	S^3 PMB + COAL	-1.75*
	2.0(b)	S^3 PMB + COAL	-1.20*

*COMPARED TO REFERENCE CASE EFFICIENCY CORRECTED FOR EQUIVALENT THERMODYNAMIC REGENERATION

Table 2.2-4. Base Case 2 Single Parameter Variations Fuel, MHD Flow Train, Size
(Reference Case Efficiency 43.5%)

REFERENCE	PARAMETRIC CASE	VARIATION	ΔEFF
MR COAL	2.1	I6 COAL	+0.89*
SUBSONIC	2.6	SUPERSONIC	-0.65
K SEED	2.5	Cs SEED	+1.12
(6-5) TESLA	2.166	4 KV/M	+0.89
	2.7	(8-7) TESLA	+1.24
1257 MWe	2.11	887 MWe	-0.75
	2.10	655 MWe	-1.67

*COMPARED TO 70% SLAG REJECTION CASE FOR MONTANA ROSEBUD (2.0(s)).

Table 2.2-5. Base Case 2 Single Parameter Variations
High Temperature Air Heater System
(Reference Case Efficiency 43.5%)

REFERENCE	PARAMETRIC CASE	VARIATION	ΔEFF
600 F AIR IN	2.16(a)	1300 F AIR IN	+0.88
2 STAGE CYCLONE COMBUSTOR	2.17	CAPFB COMBUSTOR	-1.12
600 F AIR, 900 F FLUE GAS TO PREHEAT COMBUSTOR	2.18	1300 F AIR AND FLUE GAS	+0.58
PRESSURIZED REHEAT	2.12	ATMOSPHERIC REHEAT	-0.66

two-stage cyclone with 85% slag rejection produce nearly identical calculated plant efficiencies and there is little to choose between the two design approaches. Other factors such as scalability, reliability and flow uniformity, as determined by development testing will influence the final choice.

2.2.3.2 System Results for the S^3PMB Gasifier

The two configurations using a gasifier for the first stage of the MHD combustor, e.g., the S^3PMB + coal and the S^3PMB + SPMB, were not compared directly to the reference case. The direct comparison is not correct because the gasifier configurations have a significant amount of thermal regeneration, in that they use 1300 F air which is preheated in the HRSR subsystem and the reference case has no regeneration. However, Cases 2.18 and 2.16(a) differ from Case 2.0 primarily in the degree of thermodynamic regeneration and therefore, provide a direct quantitative measure of the effect of regeneration on plant efficiency, Figure 2.2-5. To obtain a common basis of comparison, the efficiency of the reference configuration with a degree of regeneration equal to that of each of the gasifier cases was estimated using Figure 2.2-5.

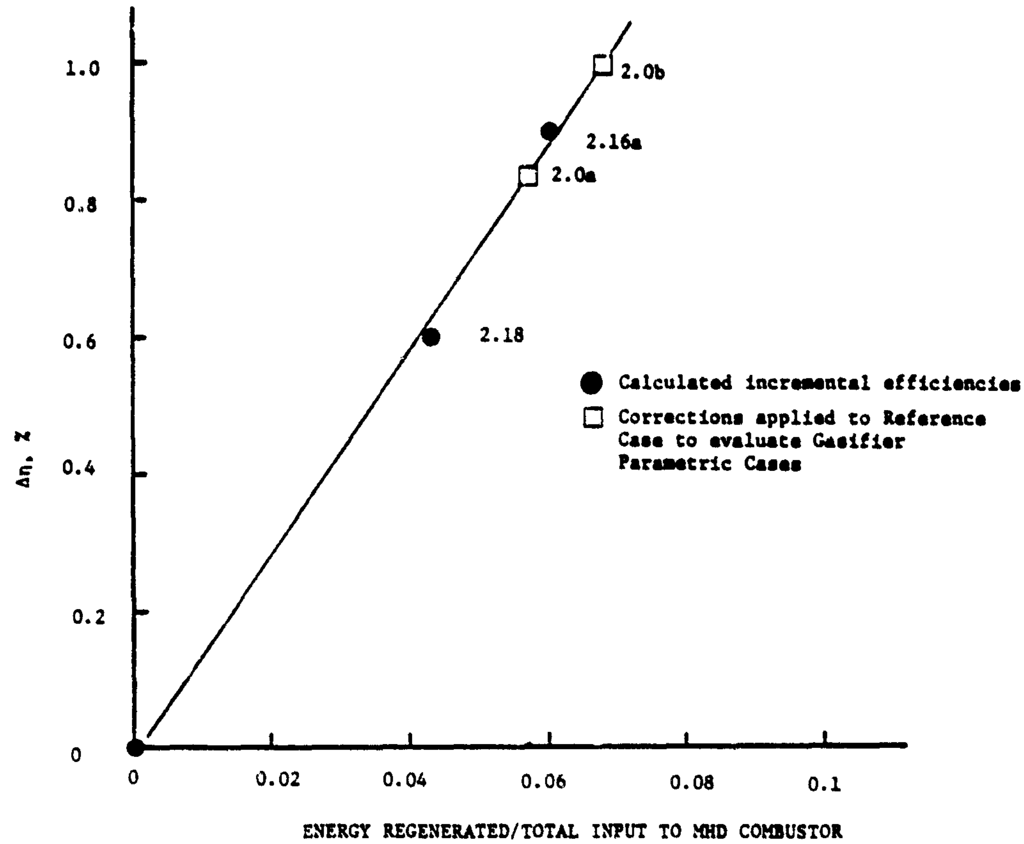


Figure 2.2-5. Effect of Regeneration on Plant Efficiency

The comparatively large performance penalty for the gasifier cases is a result of limited air preheat temperatures and high coal moisture level (see Figure 2.2-3) and heat input directly to the steam cycle (thus bypassing this energy around the MHD channel). Case 2.0(b) (S^3PMB + SPMB) has better performance than Case 2.0(a) (S^3PMB + coal) because its second-stage (3000 F) air is a smaller fraction of the total air requirement for the MHD combustor. It thus requires a smaller fuel input to the HTAH combustion system and therefore has a smaller indirectly-fired HTAH penalty.

2.2.3.3 Montana Rosebud vs Illinois #6 Coal

Using Illinois #6 coal as the fuel, rather than Montana Rosebud, increases the plant efficiency (without seed reprocessing) by 1.36 points. This improvement is a result of two major effects. First, channel performance is improved by the lower moisture content and additional heating value. Second, the significantly lower coal moisture content decreases thermal losses (see Figure 2.2-3) for evaporation in the coal dryers and in the combustion stream. When the increased seed reprocessing penalty for Illinois #6 is accounted for (see Section 3.9), the net overall efficiency increase for Illinois #6 is 0.89 points.

2.2.3.4 Channel, Magnet, and Plant Size

The next group of parametric variations, Table 2.2-4, deal with variations in channel and magnet design. Here the plant configuration is the same as the reference case, and the only variation is fractional enthalpy extraction in the channel. The discussion of the channel performance for these variations is presented in Section 3.3. The effect on plant efficiency for these cases is shown in Figure 2.2-6. There is an essentially linear relationship with a 1 percentage point increase in channel enthalpy extraction corresponding to an increase in plant efficiency of 0.48 percentage points. This quantitative relationship was determined for the reference plant configuration, but it is probably representative of all plants with an indirectly-fired HTAH subsystem.

2.2.3.5 HTAH Subsystem Configurations

In the Base Case 2 configuration with an indirectly-fired HTAH, thermal energy may be regenerated from the MHD flow train with either of two oxidizer flows, (a) the MHD combustor air, or (b) the preheat combustor air and recirculated flue gas. The regeneration is accomplished by transferring heat from the MHD flow to the oxidizer in metallic tube panels located within the HRSR subsystem. These two options for regeneration were evaluated in Cases 2.16(a) and 2.18, respectively, where (in both cases) the oxidizer was preheated to 1300 F. The rate of energy regeneration is

$$\dot{E}_R = (\dot{m}\Delta h)_{\text{oxidizer}},$$

where m is the mass flow of oxidizer through the metallic air heater and Δh is the corresponding enthalpy addition to the oxidizer. Since Δh is about equal in both cases and Case 2.16(a) uses a larger oxidizer flow in the metallic air heater than Case 2.18, the former configuration is capable of more regeneration than the latter. This result is illustrated in Figure 2.2-5.

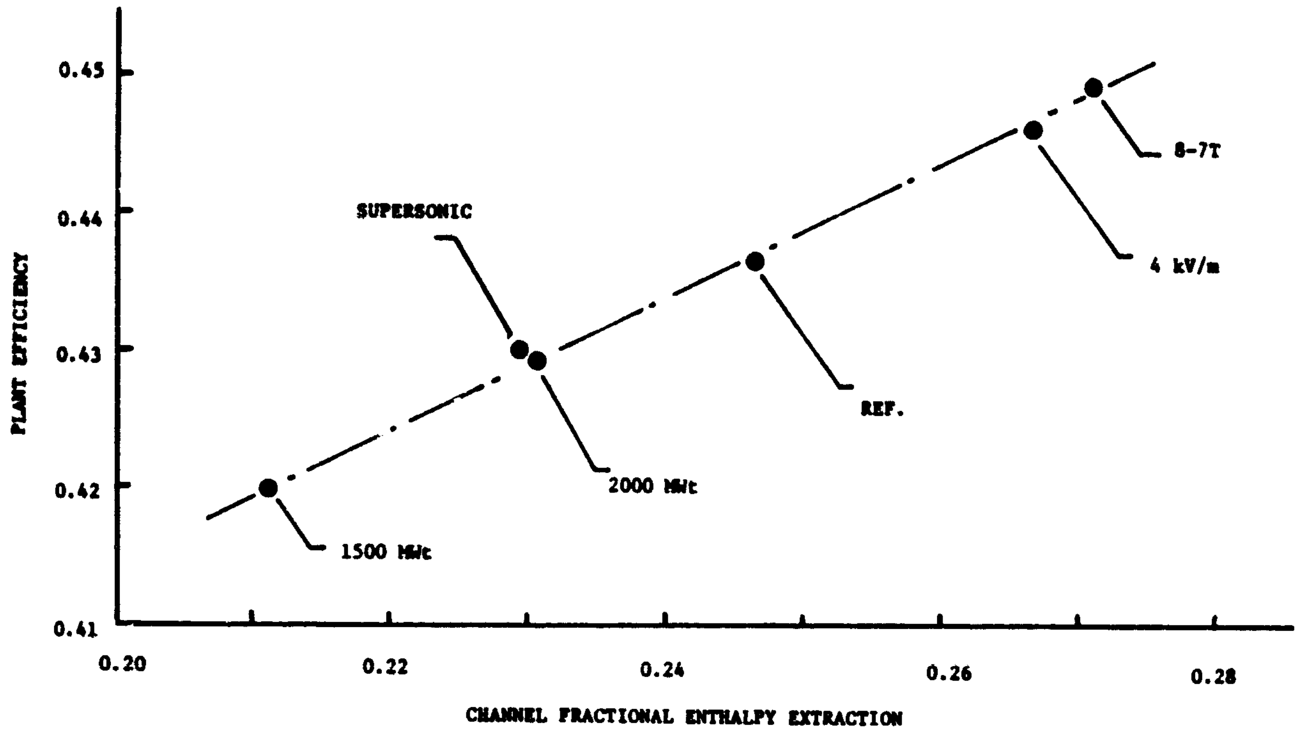


Figure 2.2-6. Effect of Channel Performance on Plant Efficiency, Base Case 2

The atmospheric HTAH combustion system (Case 2.12) resulted in a 0.66 percentage point decrease in plant efficiency as compared to the pressurized combustion system because less efficient use can be made of the flue gas exiting from the HTAH. This flue gas represents a large flow of fairly low grade heat (~ 900 F). The gas turbine, used in the pressurized system to expand the HTAH exhaust flow to atmospheric pressure, is a very efficient device in this temperature range. The gas turbine converts this low-grade energy to shaft output (which is used to drive the combustion flow) at over 50% efficiency. In contrast, the flow in the atmospheric combustion system must be driven with an electrically-powered blower (requiring ~ 10 MWe), and the low grade heat from the HTAH exhaust must be integrated into the plant. Note, Figure 2.2-3, that Case 2.12 has an above-average heat loss, as a result of the HTAH atmospheric combustion flow. In general, the pressurized HTAH combustion system is a highly efficient plant design option, in addition to the benefit of reducing component size and cost.

The use of the CAPFB combustion system for the HTAH (Case 2.17) reduced plant efficiency by 1.13 points. This degradation in performance is a result of larger thermal losses than the two-stage cyclone both to ambient and directly to the steam cycle.

2.3 BASE CASE 3

2.3.1 REFERENCE SYSTEM DESCRIPTION

The plant arrangement and state points for the reference system, Case 3.0, are given in Figure 2.3-1 and Table 2.3-1 respectively. The MHD flow train for Base Case 3 is similar to Case 2.0 in plant component arrangement except for the elimination of the indirectly-fired HTAH Subsystem and the insertion into the component lineup of two recuperative air heaters AH1 and AH2.

Table 2.3-1. State Points, Case 3.0

LOCATION	T (K)/(F)	PSIA	M (Kg/SEC)	E (MW)
1	220/373	-	-	-
2	220/373	-	-	-
3	978/1300	137.2	378.1	279.0
4	2816/4790	136.7	486.4	2608.0
5	2351/3737	17.7	486.4	1894.0
6	1866/2899	14.65	486.4	1488.6
7	1330/1935	14.54	609.7	897.1
8	692/785	14.36	609.7	400.8
9	521/478	14.22	609.7	282.
10	528/491	14.22	285.7	134.5
11	384/231	14.75	590.7*	191.2
14	608/634	150.4	378.1	124.4
21	361/190	157.	538.4	-
12	401/261	137.	538.4	-
22	423/301	400.	562.7	-
23	492/427	3860.	562.7	-
25	606/631	3670.	562.7	-
26	653/715	3610	562.7	-
27	811.1/1000	3500.	562.7	-
28	591/604	768.	552.5	-
29	811/1000	691.	557.3	-
30	314/106	2.3 "Hg	421.6	-
31	-	-	-	560.1
32	-	-	-	654.6

*Excludes Coal Drying Moisture

Case 3.0 was set up to make a comparison of the use of O_2 enriched air heated in metallic recuperative air preheaters with unenriched air heated in indirectly-fired regenerative heaters.

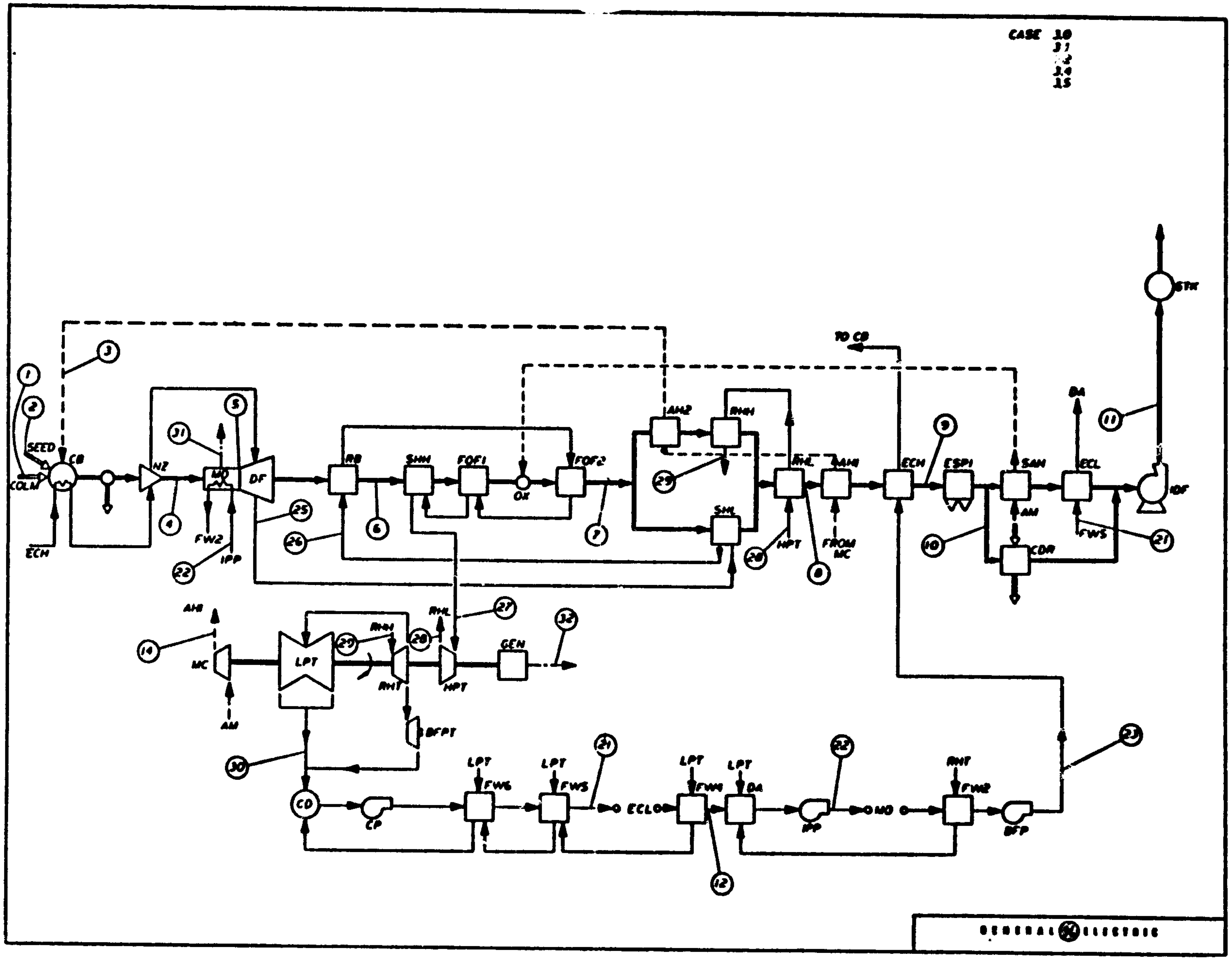


Figure 2.3-1. Plant Arrangement - Base Case 3

The practical temperature limit to which a pressurized oxidizer can be heated in a directly-fired, recuperative heater is directly related to the corrosion resistance of the materials at high metal temperatures. This limit, using available stainless steels and superalloys, is generaliy quoted at 1100 F to 1400 F. Therefore, 1300 F was chosen as the base case oxidizer preheat temperature. All of the cases in Base Case 3 assume the use of 40% additional oxygen as defined in Section 1.3. This level of enrichment was chosen based on channel performance and plant efficiency before total plant costs, including cost of the oxygen plant, were available, and a lower percentage of added oxygen may produce a lower cost of electricity (COE) by reducing the size of the required oxygen plant. This question should be examined in future studies. Also, it is assumed that the added oxygen is mixed with the incoming air at ambient pressure and temperature, although with better definition of the oxygen plant such might not be the case.

The total oxidizer stream is compressed in a non-intercooled compressor by a steam turbine driven axial flow compressor before passing through a low temperature air heater AH1, and then through the high temperature air heater, AH2. The oxidizer stream (in Case 3.0) exits AH2 at 1300 F.

Note from the plant arrangement, Figure 2.3-1, that the high temperature air heater, AH2, is part of the convection pass of the HRSR subsystem. The low temperature air heater, AH1, is located in the back pass between the low temperature reheater, RHL, and the high temperature economizer, ECH.

The hot gas for coal drying is extracted at the exit of the electrostatic precipitator which removes the solids (seed/slag) which were not collected in the upstream components of the HRSR subsystem.

The steam cycle arrangement is exactly the same as in Case 1.0. It uses two low temperature feed water heaters, FW6 and FW5, the low temperature economizer, ECL, a medium temperature feedwater heater and a deaerator, FW4 and DA, the intermediate pressure pump and the channel cooling system, IPP and MD, followed by one high temperature feed water heater and the boiler feed pump, FW2 and BFP.

2.3.2 PARAMETRIC VARIATIONS

Parametric variations are summarized in Table 2.3-2 and are self-explanatory. See Appendix B for a tabulation of efficiencies with and without seed reprocessing for each of these cases.

2.3.3 RESULTS AND CONCLUSIONS

On a plot of overall efficiency versus the ratio of channel gross output to total coal input, Figure 2.2-4, the directly-fired, recuperatively heated cases are in line with the 3000 F indirectly-fired systems using pressurized reheat combustors (Base Case 2). Thus the shaft power penalty associated with manufacturing oxygen is just counterbalanced by the avoidance of the thermal efficiencies inherent in the indirectly-fired air heater subsystem.

Table 2.3-2. Base Case 3 Single Parameter Variations
(Reference Case Efficiency 42.9%)

REFERENCE	PARAMETRIC CASE	VARIATION	ΔEFF
MR COAL	3.1	I6 COAL	+0.47
SINGLE-STAGE VORTEX COMBUSTOR	3.2	2-STAGE CYCLONE	-0.02
1300 F AIR	3.4	1100 F AIR	-0.36
(6-5) TESLA	3.5	(8-7) TESLA	+1.17

This is in contrast to Base Case 1 where a slight net loss resulted from use of an O_2-enriched flow with an indirectly-fired HTAH subsystem. Of course, these conclusions are dependent on the actual power requirement of the oxygen plant as discussed in Section 3.2.3.

Sensitivity to preheat temperature Case 3.4 is surprisingly low and it may prove desirable to lower the air preheat temperature, perhaps to 1200 F, for a more conservative design. The 100-degree drop would cost less than 0.2 percentage points.

The most important conclusion of this analysis is that competitive system efficiencies can be achieved without the necessity of a development program on regenerative heat exchangers for early commercial application.

SECTION 3

MAJOR SUBSYSTEMS AND COMPONENTS

SECTION 3

MAJOR SUBSYSTEMS AND COMPONENTS

Combustor and generator performance are clearly central to the viability of open cycle MHD power generation. Consequently a substantial effort was devoted to exploring a variety of combustor concepts, including gasification. Gasification offered the potential advantages of slag free operation, such as electrical isolation between combustor stages, and the possibility of utilizing development work already in progress. However, integration proved difficult and efficiency was relatively low. Thus, some form of close-coupled low-to-moderate slag combustion system with cyclone or vortex separation appears more promising.

Techniques for computing MHD generator performance were refined to compute optimum net power while satisfying external constraints. A close match with NASA LeRC specified generator performance was obtained. Air heater analysis was based on detailed studies done for ETF and for Base Case 3, data for O_2 plant performances was supplied by NASA LeRC.

Other subsystems covered in this section are the magnet, inverters, diffuser, heat recovery/ seed recovery, seed reprocessing and steam plants. Magnet work was done by GE's Energy Systems Products Department. Hooker Chemical Company analyzed seed reprocessing techniques and supplied general data for application to specific cases.

3.1 COMBUSTORS/GASIFIERS

3.1.1 CONCEPTS INVESTIGATED

All combustors considered in this study were nominally capable of at least 70% slag rejection. They included several varieties of the cyclone or vortex type plus a number of gasifiers. The latter offered the advantages of firing with a clean fuel and of utilizing development work already in progress but difficulties with system integration and overall efficiency resulted in a shift in emphasis to cyclones. The combustors and gasifiers considered for both the main MHD combustion system and for the air heater combustion system are described in this section since there is considerable overlap in the concepts despite the substantial difference in combustor exit temperature.

Concepts which were evaluated as fuel gas sources or heat sources for the HTAH assembly were:

1. Conventional Fixed Bed Gasifier (Wellman-Galusha)
2. Staged, High Slag Rejection Cyclone Gasifier
3. Split-Stream Slagging Pressurized Moving Bed Gasifier (S^3PMB)
4. Regeneratively Air Cooled Cyclone Coal Combustor
5. Chemically Active Pressurized Fluidized Bed Gasifier (CAPFB)

A case breakdown indicating which type of HTAH gasifier/combustor was used for each system configuration is given in Table 3.1-1.

MHD gasifiers and combustor concepts which were evaluated were:

1. Single Stage Vortex Combustor
2. Two-stage Cyclone Combustion System
3. Slagging Pressurized Moving Bed Gasifier (SPMB)
4. Split-Stream Slagging Pressurized Moving Bed Gasifier (S^3PMB)
5. S^3PMB/SPMB - the combined use of a split-stream slagging pressurized moving bed gasifier (S^3PMB) and a slagging pressurized moving bed gasifier (SPMB)
6. The combined use of a S^3PMB gasifier with a single stage suspension type pulverized coal combustor

A case breakdown for the use of these combustor systems in the PSPEC study is given in Table 3.1-2.

3.1.2 COMPONENT DESCRIPTIONS

Brief descriptions of the design analysis used and the data available on these components are presented below.

3.1.2.1 Wellman Galusha Fixed Bed Gasifier

This gasifier was used as a fuel gas source for the HTAH assembly in Base Case 1. It was specified to be an off-the-shelf gasifier operating at atmospheric pressure. Several conventional designs[1] were examined, Table 3.1-3, and the Wellman Galusha fixed bed gasifier, Figure 3.1-1, delivering hot raw gas to the HTAH assembly, was selected after consultation with NASA personnel.

Performance data were obtained from the manufacturer[2]. With a stirring rod, this design has demonstrated operational capability with both caking and moderately caking coals, thus it is suitable for both Montana Rosebud (non-caking) and Illinois #6 (caking) coals. Capacity of a 10 ft. I.D. vessel while operating with 'as received' Montana Rosebud is estimated by the gasifier manufacturer to be 1.070 Kg/s, and with Illinois #6 is about 0.945 Kg/s. Other performance data for the Wellman-Galusha are summarized in Table 3.1-4.

A process flow diagram with detailed state point data is included in the discussion of the HTAH subsystem. In scaling to commercial size plant, banks of 10 ft. I.D. gasifiers were utilized, the number of gasifiers being determined by the thermal requirements of the HTAH assembly. For all the cases considered, 'as received' coal was the fuel for the gasifiers.

Table 3.1-1. Case Breakdown for HTAH Gasifiers/Combustors

Date: 9 February 1979
Revision #1

Type Gasifier or Combustor	Case #	Coal Type	As Fired FMR	T_{air} (in) ^oF	$T_{Product}$ (out) ^oF	$\dot{m}_{Coal}$ Kg/s (as fired)	Remarks
1. Conventional Fixed Bed Gasifiers - Data for Wellman-Galusha, 10 ft ID gasifier, atmospheric pressure operation, hot raw gas fed to HTAH assembly for combustion of low Btu gas.	1.0	MR	29.37	77	730	37.01	SOA Base Case w/MR
	1.1	I6	9.77	77	1029	28.53	SOA Base Case w/I6
	1.2	MR	29.37	77	730	44.53	Air only
	1.3	MR	29.37	77	730	37.01	Same as 1.0
	1.4(a)	MR	29.37	77	730	37.01	Same as 1.0
	1.4(b)	I6	9.77	77	1029	28.53	95% slag rejection for MHD combustor
2. Two-Stage Cyclone Combustor - Design data from GE ETF study, 85% slag rej[illegible]	2.0	MR	5.0	648	2750	42.19	Base Case
	2.1	I6	2.0	649	2750	37.42	Estimated $\dot{m}_{Coal}$ from Cases 1.0 and 1.1
	2.2	MR	5.0	648	2730	32.4	Hot Bottom HTAH, impacts gasifier scaling
	2.3	--	--	--	--	--	
	2.4	MR	5.0	648	2750	41.33	Same as 2.0
	2.5	MR	5.0	648	2750	41.56	Same as 2.0, Cs seed
	2.6	MR	5.0	648	2750	43.71	Same as 2.0, supersonic channel
	2.7	MR	5.0	648	2750	42.19	Same as 2.0, 8T magnet
	2.8	--	--	--	--	--	
	2.9	--	--	--	--	--	
	2.10	MR	5.0	648	2750	23.27	1500 MWt to main combustor
	2.11	MR	5.0	648	2750	30.58	2000 MWt to main combustor
	2.13	--	--	--	--	--	
	2.15	MR	5.0	648	2750	42.19	Same as 2.0, dual MHD combustor and channel flow train
	2.16	MR	5.0	1100	2750	32.99	Hot Bottom HTAH, impacts gasifier scaling
3. Top Gas from S^3PMB - Advanced concept for slagging moving bed gasifier, clear gas to MHD flow train	2.0(a)	MR	12.93/5.0	1100	706	68.88/42.58*	S^3PMB plus coal (matched to HTAH and MHD combustor)
	2.0(a-1)	MR	12.93	1300	706		S^3PMB plus parallel plant (matched to MHD combustor)
	2.0(b)	MR	12.93/12.93	1100	706	52.29/63.22**	S^3PMB plus SPMB (matched to HTAH and MHD combustor)
	2.1(a)	I6	2.0	1300	710	--	S^3PMB plus parallel plant (matched to MHD combustor)
	2.1(b)	I6	2.0	1300	710	--	S^3PMB plus SPMB (matched to HTAH and MHD combustor)
4. Regeneratively Air Cooled Cyclone Combustor - Combustors mounted in HTAH vessels.	2.12	MR	5.0	1100	3300	40.65	5 burners per HTAH vessel
5. CAPFB - Chemically Active Pressurized Fluidized Bed Gasifier	2.17	MR	12.93	404/818+	1650	51.21	Except for HTAH gasifier, same as Case 2.0

+404 F to CAPFB
818 to HTAH combustor

*68.88 Kg/s to S^3PMB/42.58 Kg/s to second stage combustor
**52.2 Kg/s to S^3PMB/63.22 Kg/s to SPMB

Table 3.1-2. Case Breakdown for MHD Combustors

Type Gasifier or Combustor	Case #	Coal Type	As Fired FMR	T_{air} (In) °F	$T_{product}$ (Out) °F	$\dot{m}_{coal}$ Kg/s (as fired)	Remarks
1. Single Stage Vortex Combustor, 70% slag rejection, 5% ΔP	1.4	MR	5.0	2700	4600$^+$	79.40	$^+$Nominal temperature, also 85% slag rejection
	1.5	I6	2.0	2700		72.18	Coal change impact
	2.0	MR	5.0	3000		70.36	Base Case, also 85% slag rejection
	2.1	I6	2.0	3000		63.98	Estimated from cases 1.0 and 1.1
	2.3	--	--	--		--	
	2.5	MR	5.0	3000		70.36	Cs seed, same as 2.0
	2.6	MR	5.0	3000		70.36	Supersonic channel, same as 2.0
	2.7	MR	5.0	3000		70.36	8T magnet, same as 2.0
	2.8						
	2.9						
	2.10	MR	5.0	3000		37.96	1500 MWt to main combustor
	2.11	MR	5.0	3000		50.26	2000 MWt to main combustor, also 85% slag rejection
	2.12	MR	5.0	3000		70.36	Regeneratively air cooled combustor for HX, same as 2.0
	2.13						
	2.14	MR	5.0	3000		TBD	Directly fired
	2.15	MR	5.0	3000		70.36	Dual combustor and channel flow trains
	2.16	MR	5.0	3000		70.36	Hot bottom HX
	2.17	MR	5.0	3000		70.36	CAPFB for HX
	3.0	MR	5.0	1300		98.32	Base Case, recuperative air heat, O_2 enrichment
	3.1	I6	2.0	1300		89.36	TBD
	3.4	MR	5.0	1100		100.00	1100 F air + O_2
	3.5	MR	5.0	1300		98.32	Same as Case 3.0
2. Two Stage Cyclone Combustor, ≥ 85% slag rejection, 10% ΔP	1.0	MR	5.0	2700		79.40	Base Case, air/O_2 blown
	1.1	I6	2.0	2700		72.18	Coal change impact
	1.2	MR	5.0	2700		73.23	Air blown
	1.3	MR	5.0	2700		79.40	Same as 1.0
	2.2	MR	5.0	3000		70.36	Hot bottom HX, no impact on combustor
	2.4	MR	5.0	3000		70.36	Staged cyclone replaces single stage vortex
	3.2	MR	5.0	1300		98.32	Same as case 3.0

Table 3.1-2. Case Breakdown for MHD Combustors (Continued)

Type Gasifier or Combustor	Case #	Coal Type	As Fired PMR	T_{air} (in) °F	$T_{product}$ (out) °F	$\dot{m}_{coal}$ Kg/s (as fired)	Remarks
3. Advanced Slagging Moving Bed Gasifier (SPMB)	3.0(a)	MR	12.93	1300	2524	102.16	Deleted after analysis
	3.1(a)	I6	2.0	1300	--	--	Deleted after partial analysis
4. Split-Stream Slagging Pressurized Moving Bed Gasifier (S^3PMB)	2.0(a-1)	MR	12.93	1300	706/3340	176.0	Deleted after analysis
	2.1(a)	I6	2.0	1300	--	--	Deleted after partial analysis
5. S^3PMB/SPMB	2.0(b)	MR	12.93	1300/1300	706/3340/2535	52.27/63.22	
	2.1(b)	I6	2.0	1300/1300	--	--	Deleted after partial analysis
6. S^3PMB/Suspension Type Pulverized Coal Combustor	2.0(a)	MR		1300/[illegible]00	706/3340/4600	68.88/42.58	New case to replace 2.0(a-1), 38% and 91% slag rejection

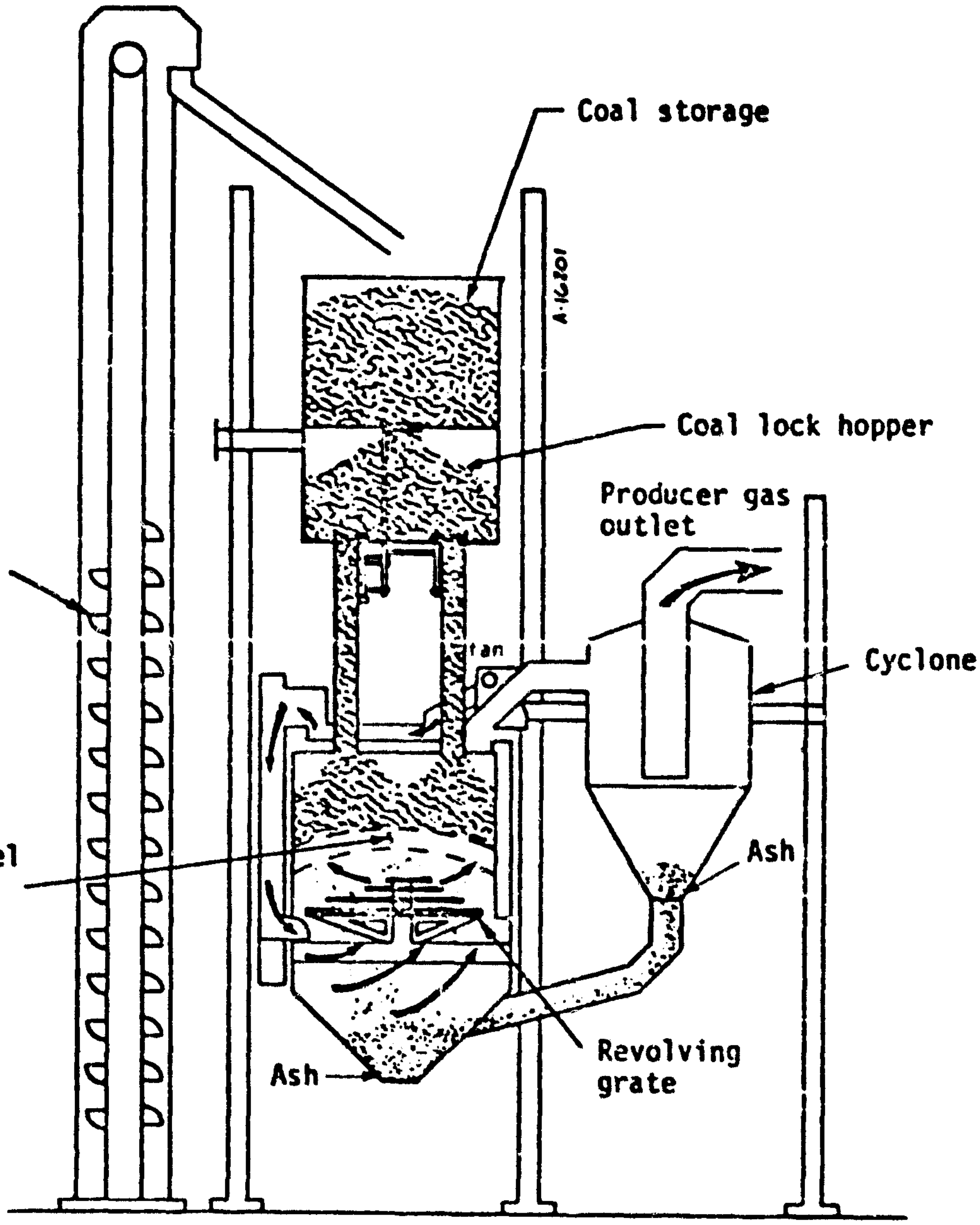


Figure 3.1-1. Wellman-Galusha Gasifier

Table 3.1-3. Gasifiers Considered for HTAH Application in Base Case 1

	NOTES	BED TYPE	NUMBER STAGES
LURGI	1	FIXED	1
WELLMAN-GALUSHA		FIXED	1
WILPUTTE GASIFIER		FIXED	1
RILEY-MORGAN		FIXED	1
WELLMAN		FIXED	2
STOIC (FOSTER-WHEELER)		FIXED	2
WOODALL-DUCKHAM		FIXED	2
KOPPERS-TOTZEK	2	ENTRAINED	1
WINKLER		FLUIDIZED	1

NOTES: 1. ONLY PRESSURIZED TYPE

2. O_2 ONLY, ALL OTHERS AIR OR O_2

3.1.2.2 Two-Stage Cyclone Gasifier/Combustor

The high slag rejection cyclone gasifiers and combustors used in this study were extensions of the GE ETF[3] high slag rejection gasifier/combustor, Figure 3.1-2. The first stage consists of a cluster of vertical primary vortex gasifiers which feed into a common cyclone slag separator-receiver. The fuel gas from the first stage fires a second stage gas phase combustor which may be integral with or separated from the first stage gasifier. When used as an MHD combustor, the first stage gasifiers are manifolded to a single second stage gas phase combustor. The gasifier is designed for 85% slag rejection with an 8% pressure drop. The gas phase combustor has a 2% pressure drop. Heat and mass balances were determined from equilibrium analysis. At nominal design conditions, the first stage wall heat transfer is 5%, and the second stage wall heat transfer is 2% of the input coal HHV.

For both the first and second stages, 98% of the wall heat transfer was assumed recovered by cooling water, with the remaining 2% lost to ambient. For conditions perturbed from the reference design point, wall heat transfer rates were scaled as a function of chamber pressure, chamber temperature and combustion product mass flow rate.

As a gasifier for the HTAH subsystem the high slag rejection cyclone concept is modified as follows:

1. The second stage gas phase combustor shown in Figure 3.1-2 will not be used. Instead, a connecting duct is manifolded to the HTAH pressure vessels and final combustion of the fuel gas occurs in the combustion domes for each HTAH vessel.

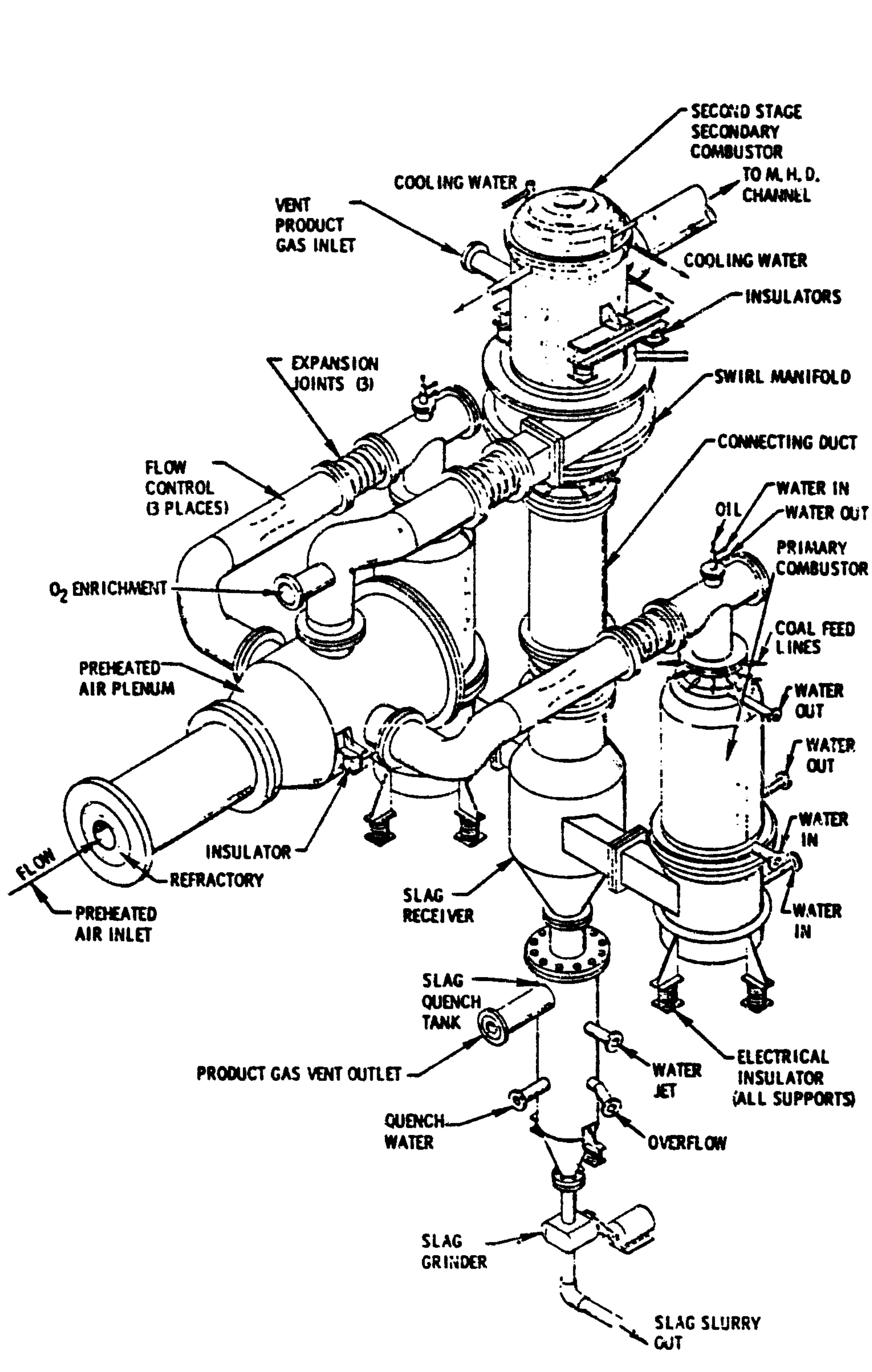


Figure 3.1-2. Two-Stage High Slag Rejection Cyclone Gasifier/Combustor

Table 3.1-4. Nominal Operating and Performance Data - A Wellman Galusha Gasifier

	MONTANA ROSEBUD	ILLINOIS #6
GASIFIER DIAMETER (FT)	10	10
OPERATING PRESSURE (ATM)	1.1	1.1
CAPACITY (KG/SEC)	1.070	0.945
OPERATIONAL LIMITATION	NONE	STIRRING ROD REQUIRED
COST - VESSEL + AUXILIARIES	$300,000	$300,000
- INSTALLATION AND CONTROLS	$180,000	$180,000
SPACE ENVELOPE (FT X FT X FT) INCLUDING DUST COLLECTOR	16 x 28 x 70	16 x 28 x 70
WEIGHT (LBS)		
- EMPTY WITH CYCLONE	90,000	90,000
- COAL LOAD	50,000	50,000
HOT RAW GAS EFFICIENCY	92%	92%
RAW GAS TEMPERATURE, ^{o}F	730	1029
PRESSURE DROP, %	10	10

2. The number of gasifiers clustered around a slag receiver will depend on the thermal power requirements for the HTAH assembly; however, up to 5 primary gasifiers can be used for a given cluster. For the scale of plants investigated, clusters of between 3 to 5 gasifiers were utilized.

3. The primary HTAH gasifiers operate at lower gasification air temperatures (589 K to 867 K) than the ETF design (1757 K to 1922 K) and deliver the low Btu fuel gas at lower temperatures, nominally 1783 K versus 2088 K.

3.1.2.3 Slagging Pressurized Moving Bed Gasifier

The slagging, pressurized moving bed* gasifier (SPMB), Figure 3.1-3, is an extension of an oxygen/steam blown gasifier concept presently being developed by the Grand Forks Energy Technology Center[4,5] and the British Gas Corporation[6,7].

The Grand Forks and British gasifiers are similar in concept in that a hearth and blast furnace type tuyeres replace a conventional grate support, thus permitting slagging operation. Slagging operation relaxes the constraints on input air temperature and the steam/oxygen ratio as compared to dry ash operation. However, slag poses some significant materials problems and long term testing is required to demonstrate commercial acceptability. Throughputs are 2 to 4 times higher than that of dry ash fixed bed gasifiers.

For MHD applications, this concept was modified to air blown operation to avoid the adverse effects of OH^- on plasma conductivity. An idealized representation of the moving bed gasifier[8] is shown in Figure 3.1-4. Coal is fed in at the top of the reactor and moves down by gravity flow. At the top of the gasifier, the coal is preheated and dried through heat exchange between the coal and hot gases from below. In the drying zone, as shown in Figure 3.1-5, there are rapid changes in both gas and coal temperature primarily resulting from water vaporization. As the coal proceeds downward it enters a devolatilization zone where coal gases and tars are expelled. The temperature changes in this zone are relatively small since the heat absorbed by the coal char is nearly equal to the heat released during pyrolysis. In the third region, endothermic gasification reactions occur, and the temperature rises rapidly. In the gasification zone carbon conversion occurs primarily via the following reactions:

$$C + H_2O \rightleftarrows CO + H_2 \tag{1}$$

$$C + CO_2 \rightleftarrows 2\,CO \tag{2}$$

$$C + 2H_2 \rightleftarrows CH_4 \tag{3}$$

The gas phase water-gas shift reaction catalyzed by coal particles also plays an important role.

$$CO + H_2O \rightleftarrows CO_2 + H_2 \text{ (Water-gas shift)} \tag{4}$$

The combustion zone in the vicinity of the tuyeres produces a rapid rise in temperature with the following combustion reaction dominating:

$$rC + O_2 \rightarrow 2(r - 1)\ CO + (2 - r)\ CO_2, \tag{5}$$

where r is a system constant dependent on reaction conditions.

* Conventional Fixed bed gasifiers are actually slowly moving bed systems. Within this context, the term 'fixed bed' and 'moving bed' are often used interchangeably in coal gasification literature.

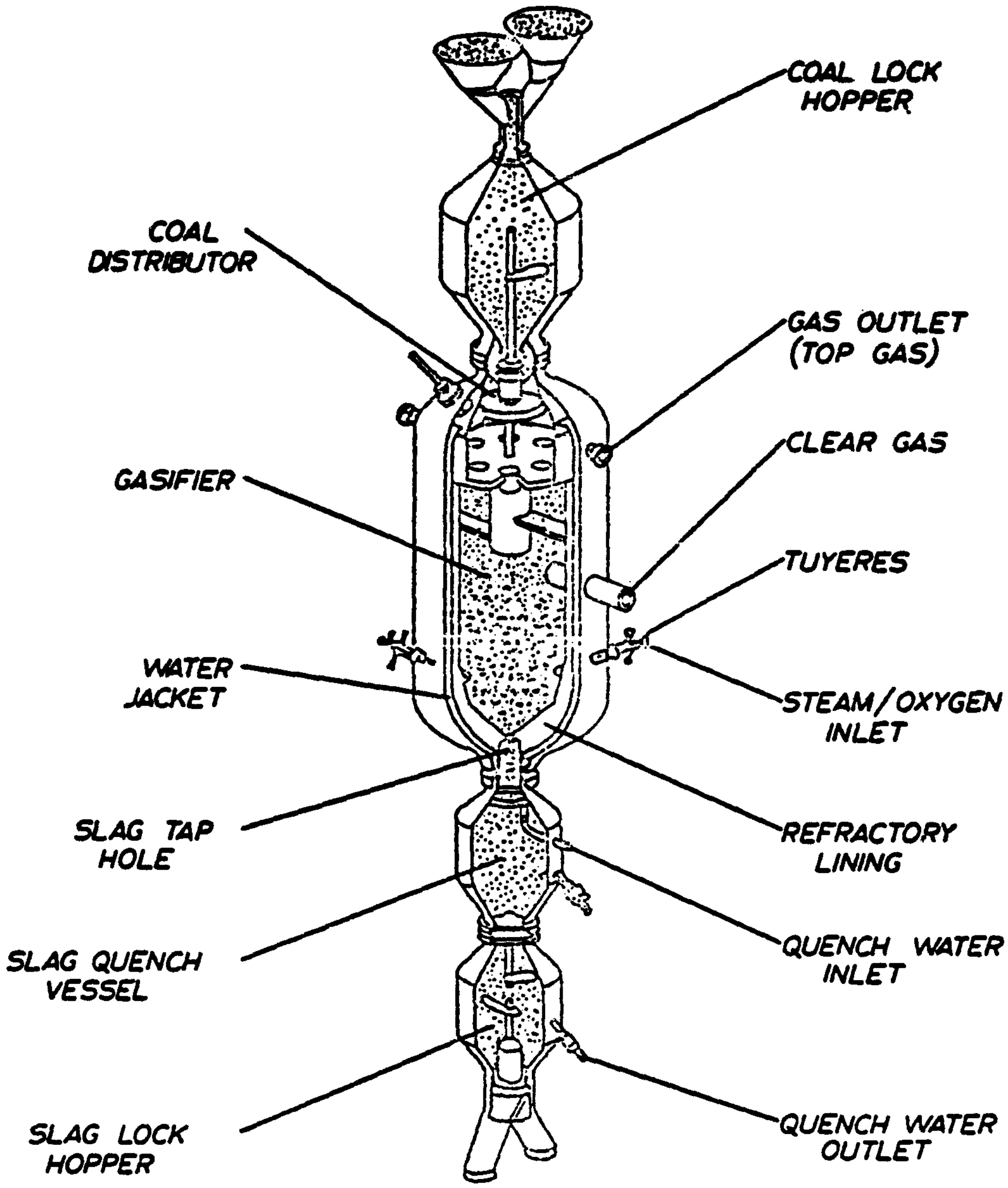


Figure 3.1-3. Slagging Pressurized Moving Bed Gasifier (Reference 8)

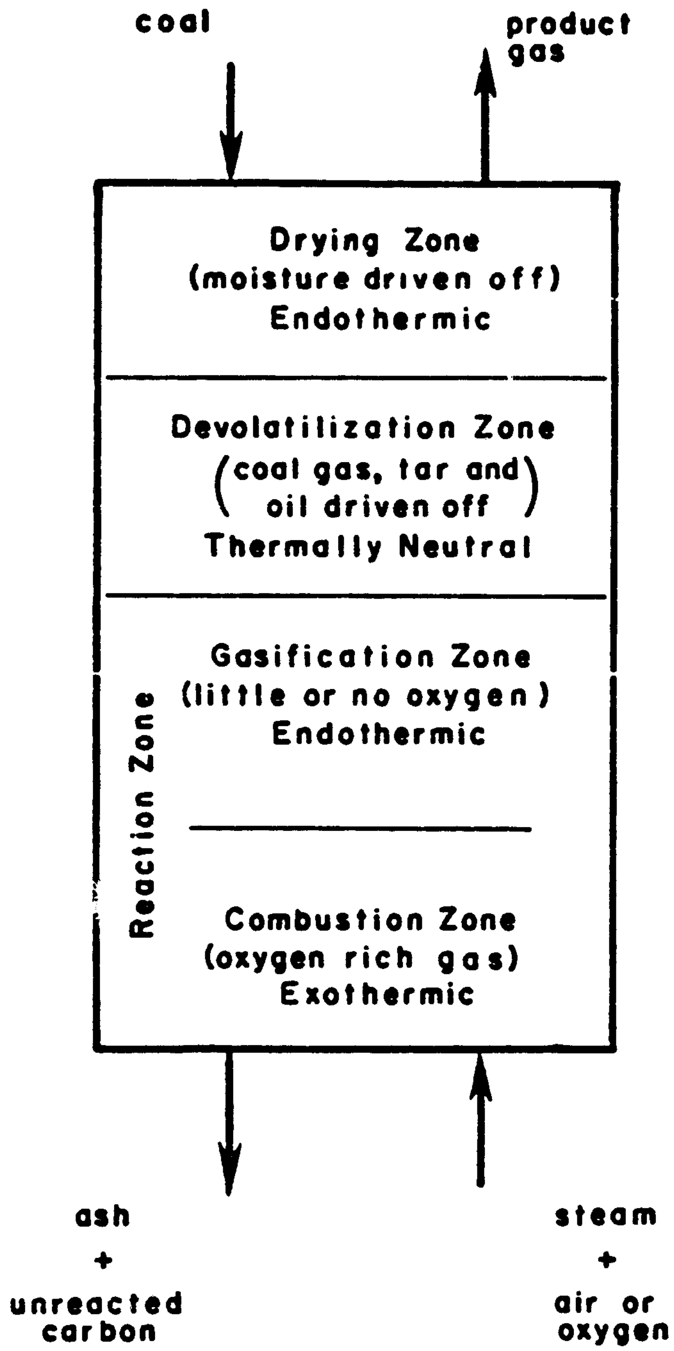


Figure 3.1-4. Schematic of Moving Bed Gasifier (Reference 8)

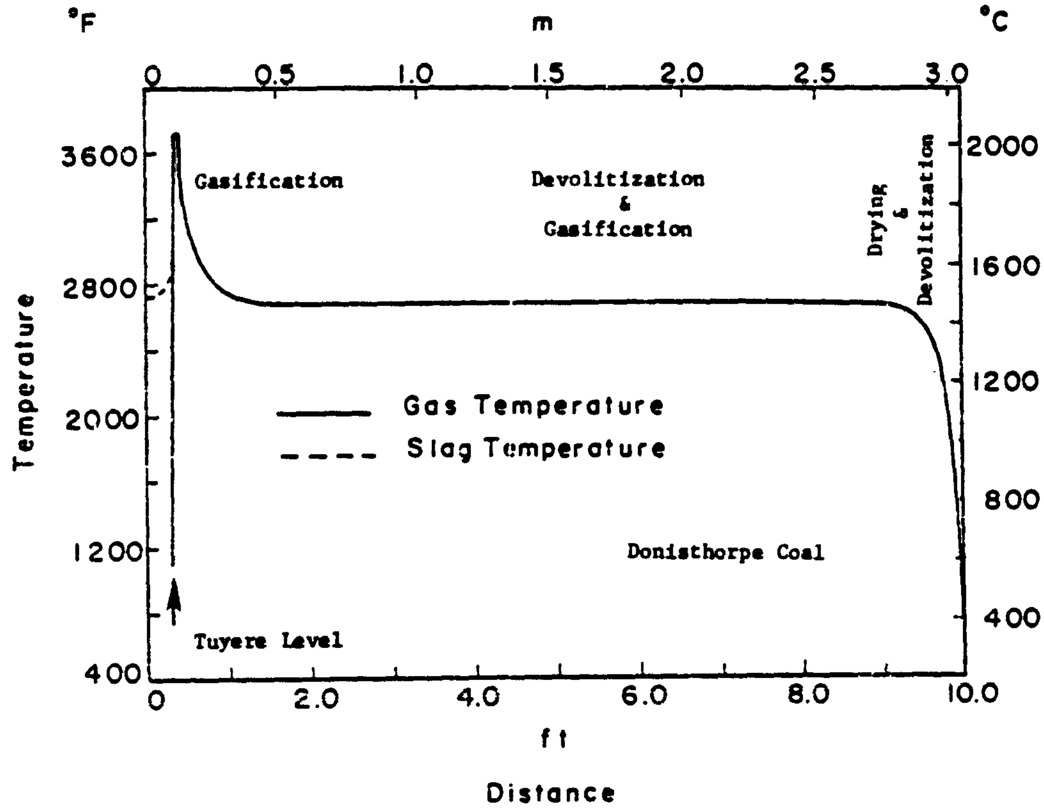


Figure 3.1-5. Predicted Temperature Profile in a Moving Bed Gasifier (Reference 8)

Although several models have been developed for analysis of fixed bed gasifiers[8,9,10], accurate prediction of fuel gas composition is difficult because of non-equilibrium effects associated with the heterogeneous reactions between the solid carbon and gaseous reactants, as well as uncertainties associated with the pyrolysis process. Empirical correlations and experimental data for the drying and devolatilization of the coal[11,12,13] were utilized for the upper region of the gasifier. An equilibrium analysis was used for the gasification and combustion of the char in the lower section of the gasifier. In the devolitization and drying regions, reactions between the coal gas and upflowing hot product gas were neglected. The end of the devolatilization zone was set by a char temperature of 1273 K, for which data indicate pyrolysis is essentially complete for most coals.

Having established the char composition, equilibrium analysis was then used to establish an appropriate stoichiometry for complete carbon conversion based upon the char composition. The calculated gas temperature and mass flow at the boundary of the combustion and gasification zones were then used to determine the fuel gas temperature exiting the gasifier by means of an energy balance. The fuel gas consists of a mixture of coal gas, tars, oils, phenols, and product gas. This segmented analysis predicts higher gasifier exit temperatures than does a total equilibrium analysis since a larger percentage of water decomposition is predicted in the latter case.

The coal delivered to the gasifier was 'surface dried' to a FMR of 12.93 for the Montana Rosebud coal and a FMR of 2.0 for the Illinois #6 coal. The gasification air temperature was limited to 1300°F* (978 K) and the fuel gas exit temperature to about 1700°F (1200 K). This exit gas temperature is in the range of operating conditions which have been demonstrated by the British Gas Corporation. Higher inlet air temperatures would lead to better MHD generator performance, but operation under these conditions is questionable in the absence of experimental data. In addition to the more severe material problems associated with higher bed temperature, there is a risk of bed agglomeration.

At nominal design conditions, the gasifier has a 3.5% wall heat transfer rate, 60% of which is transferred to cooling water and the remainder lost to the ambient air. The pressure drop across the gasifier was set at 8%. The I.D. of the gasifier was assumed to be 4 meters, which is the nominal size of the gasifier being designed and developed for combined cycle applications. At a design pressure of 7.4 atm, this size unit has nominal coal capacity of 4.71 Kg/sec with 'as fired' Montana Rosebud (FMR-12.93).

3.1.2.4 <u>Split Stream Slagging Pressurized Moving Bed Gasifier (S^3PMB)</u>

The S^3PMB is an extension of the SPMB to split-stream operation. The 'clear gas' from the gasifier, see Figure 3.1-3, is rich in CO. It is delivered to the MHD combustor where its combustion results in a very high temperature plasma nearly free of electron-absorbing OH radicals. Both effects tend to increase electrical conductivity. The 'top gas,' containing moisture and hydrocarbons is used to fire the HTAH combustors where neither extreme temperature nor high conductivity are important.

An idealized schematic of the S^3PMB is given in Figure 3.1-6. It is divided into 3 major sections. Section I is a drying and devolatilization zone and is the same as for the SPMB. Sections II and III are gasification and combustion zones for the char, where a portion of the product gas is used to devolatilize and dry the coal in Section I. The chars and product gas compositions are the same as for the SMPB.

The mass flow split of product gas determines the energy split between the top gas and clear gas. The heating value of the coal gas and tars given off in Section I are quite high and system energy requirements suggest that the mass flow of product gas for devolatilization be as small as possible. The limit for this ratio was determined by the minimum top gas temperature for suitable transport and combustion of the tar and oil vapors set at 700°F (644 K). Even at the limiting mass flow split, fuel gas to the MHD combustor was insufficient and required supplementing by additional coal or SPMB fuel gas. Changing the degree of coal drying and the level at which the clear gas is extracted from the gasifier could also alter the energy split between the HTAH and MHD combustor, but these approaches for design analysis were investigated only to a limited extent**.

* This air temperature is also compatible with recuperative air heater performance capabilities.

** By pre-drying the coal, the total energy in the top gas is reduced but not sufficiently to eliminate an energy mismatch between the HTAH amd MHD combustor. Splitting the streams higher in the gasifier will lead to practical design problems since the split would have to be in a region where rapid changes in temperature and composition occur.

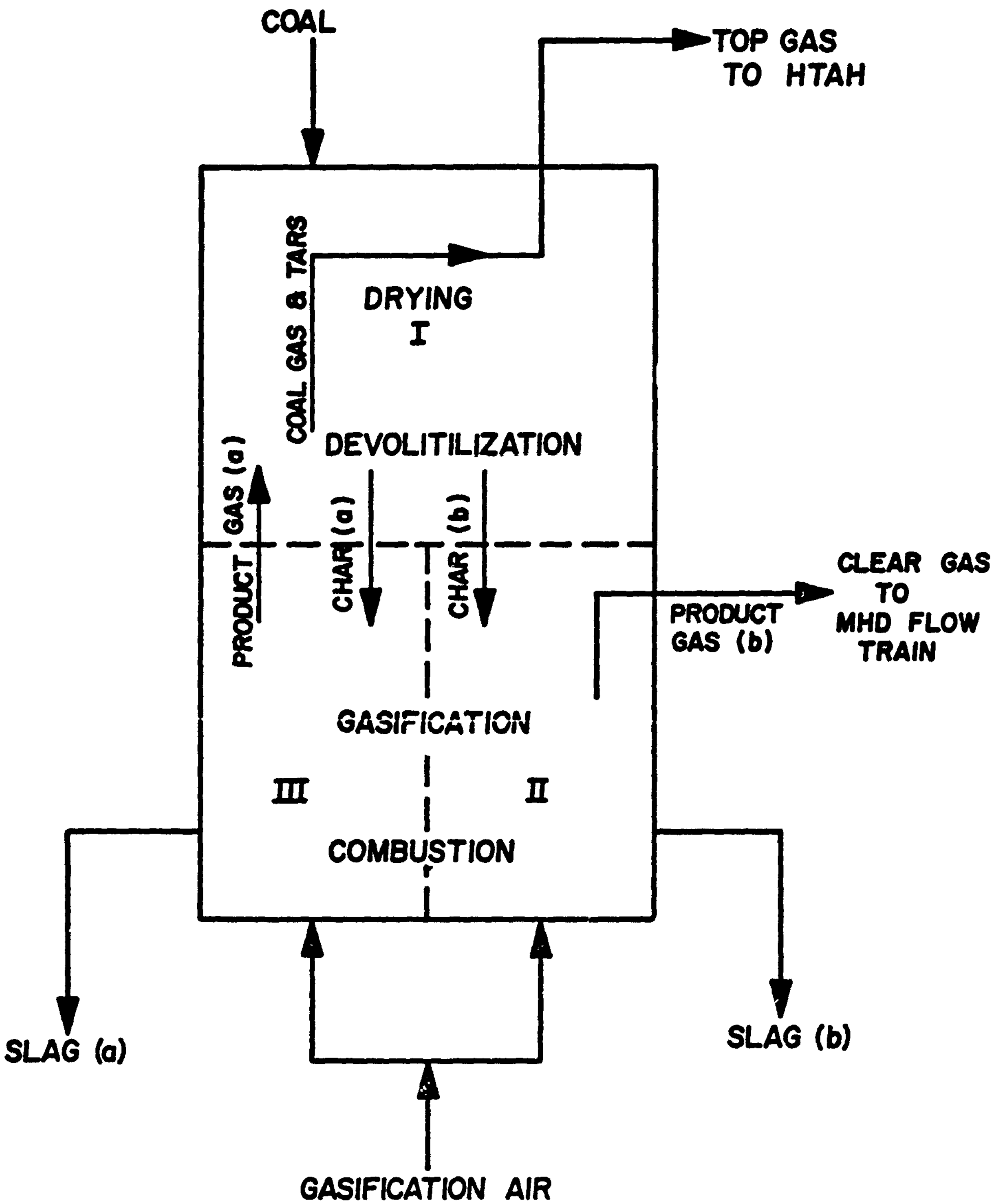


Figure 3.1-6. Schematic of Split-Stream, Slagging Pressurized Moving Bed Gasifier (S^3PMB)

To achieve an appropriate thermal balance between the HTAH assembly and the MHD combustor, supplemental fuel was used for the MHD flow train. Two possibilities were evaluated. In Case 2.0(a), pulverized coal was injected into the second stage MHD combustor to provide the appropriate energy match. MHD generator analysis was done assuming both zero and 85% slag rejection but generator performance was poor in the zero rejection case and system Case 2.0(a) was based on the 85% slag rejection assumption. In that case the effective slag rejection for the combined streams is 91%. In Case 2.0(b), the fuel gas from an SPMB gasifier was used to supplement the fuel supplied to the MHD flow train so that the combustion products are essentially slag free. Final determination of mass and energy balances required iteration with the MHD generator analysis since fuel compositions affect optimum generator performance.

3.1.2.5 Regeneratively Air-Cooled Cyclone Coal Combustor

In Case 2.12, regeneratively air cooled vortex coal combustors are the sources for the HTAH assembly. This component is an extension of work currently being performed by GE as part of the GE/DOE Closed Cycle MHD Program[14] in which a slag rejection rate of ~ 90% and a carbon conversion rate of 99% have been demonstrated while providing a combustion gas temperature of approximately 3350°F. The design provides for:

1. Integrated axial swirl injection of pulverized coal, pilot fuel, primary air and secondary air to improve gas flow field symmetry and to reduce slag induced wall geometrical effects.

2. Controllability of the slag and inner refractory interface temperature in order to establish a uniform slag layer flow which provides minimum modification of the gas flow field.

3. Regeneration of the combustor heat loss into the MHD cycle by using the combustor cooling air as secondary combustion air.

4. Positive bypassing of combustion product flow through the slag tap to provide continuous outflow of the slag without mechanical assistance.

A cut-away view of the prototype combustor is shown in Figure 3.1-7. It is mounted horizontally and has an inner diameter of 12 inches and a length of 24 inches. Injector swirl vanes are radiatively and convectively cooled and easily changeable allowing variation of the swirl angle and injection velocity. A water-cooled pilot gas burner, inside the secondary swirl vanes, is used to preheat the combustor system to prevent thermal shock and to attain refractory temperatures close to thermal equilibrium for coal combustion.

The refractory liner is made of Emerald Plastic, a commercial chrome-alumina material. The liner remains in tight contact with a combustor inner metal shell which has 72 air cooling grooves on its outer surface. To minimize heat loss from the air coolant to the environment, a castable refractory material covers the combustor vessel inner shell. Both the combustor's refractory lining and the inner metal shell expand considerably in axial and radial directions, so special high temperature metal bellows seals were designed and fabricated in order to accommodate differences of thermal expansion and to avoid gas exchanges between combustion products and cooling air.

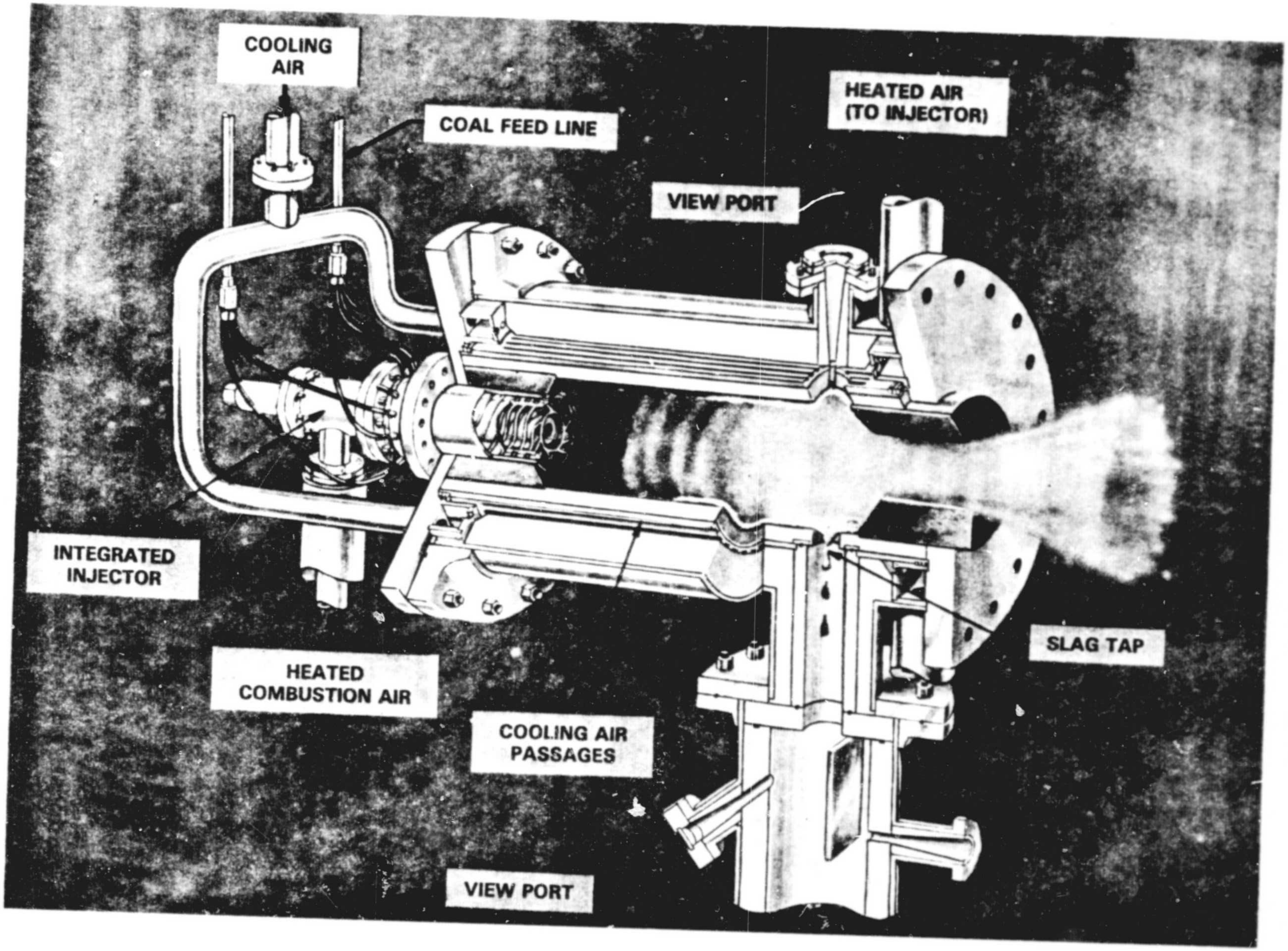


Figure 3.1-7. Regeneratively Air Cooled Cyclone Coal Combustor

For Case 2.12 each HTAH vessel will have a bank of 5 combustors. Each unit is 6' I.D. x 12' I.L., with a rated capacity of 0.45 Kg/sec. Experimental data indicates that ≥ 95% of the wall heat transfer will be recovered in the regenerative air cooling loop with the remainder lost to ambient. The wall heat transfer for the scaled combustor is 12% of the input coal HHV so that only 0.6% is lost to ambient. The pressure drop for the combustor, including the regenerative air cooling loop, is 10%.

3.1.2.6 Single-Stage Vortex Combustor

The single-stage vortex combustor is assumed to be similar in design to that proposed by Avco in the ETF Study[15]. To scale this combustor to commercial size plants, ETF size modules (4' I.D. x 13.6' I.L.) were used. At a design pressure of 6.0 atm, the combustor wall heat transfer is 5% of the 'as fired' input coal thermal energy, with 98% of this thermal energy transferred to cooling water and the remaining 2% lost to ambient. A pressure drop of 5% was used, as in the ETF study. No account was taken of heat losses in manifold ducting. Analysis was done for both 70% and 85% slag rejection.

3.1.2.7 Chemically Active Pressurized Fluidized Bed

The chemically active pressurized fluidized bed (CAPFB) was investigated as a 'clean' fuel gas source for the HTAH in Case 2.17. The performance analysis for this gasification concept was done by the Foster Wheeler Development Corporation. Data developed during the GE ETF Study was used to establish heat loss and pressure drop performance. At a design operating pressure of 10.2 atm, three gasifier vessels each 18.3' I.D. x 30' I.L. are required. The gasifier vessels, Figure 3.1-8, include a sulfate generator compartment for a once-through sulfur removal process. Each gasifier has an 'as fired' coal capacity of 17.07 Kg/sec of a pressure of 10.2 atm. About 1.4% of the coal thermal input is lost to ambient and 8% is transferred to internal water cooling tube bundles which are used to control bed temperature. An additional 0.9% heat loss is associated with spent bed material discharged from the sulfate generator.

The 'off gas' from the sulfate generator ($T = 1600^{o}F$) accounts for 4.7% of the HHV but most of this is recovered by recuperative heat exchangers as recirculated flue gas at a temperature of $800^{o}F$ (700 K). Cyclone separators collect particulates and return them to the bed. The sensible plus chemical energy of the fuel gas ($T = 1650^{o}F$) delivered to combustors mounted in the domes of the air heaters contains 85.5% of the total thermal input (coal, solid sorbents, and air) to the gasifier.

3.1.3 HEAT LOSS SCALING FOR MHD GENERATOR

System analysis of MHD power plants incorporating the variety of gasifiers/combustors considered was complicated by the manner in which these components are coupled to the high temperature air heater and the MHD generator, as well as the manner in which the combustor/gasifier heat losses scale with fuel loading, operating pressure and operating temperature. The coupling of the combustor/gasifier with the high temperature air heater and the MHD generator depends on the design constraints of the particular combustion/gasification system under evaluation (e.g., air preheat temperature limits, equivalence ratio limits, coal preparation

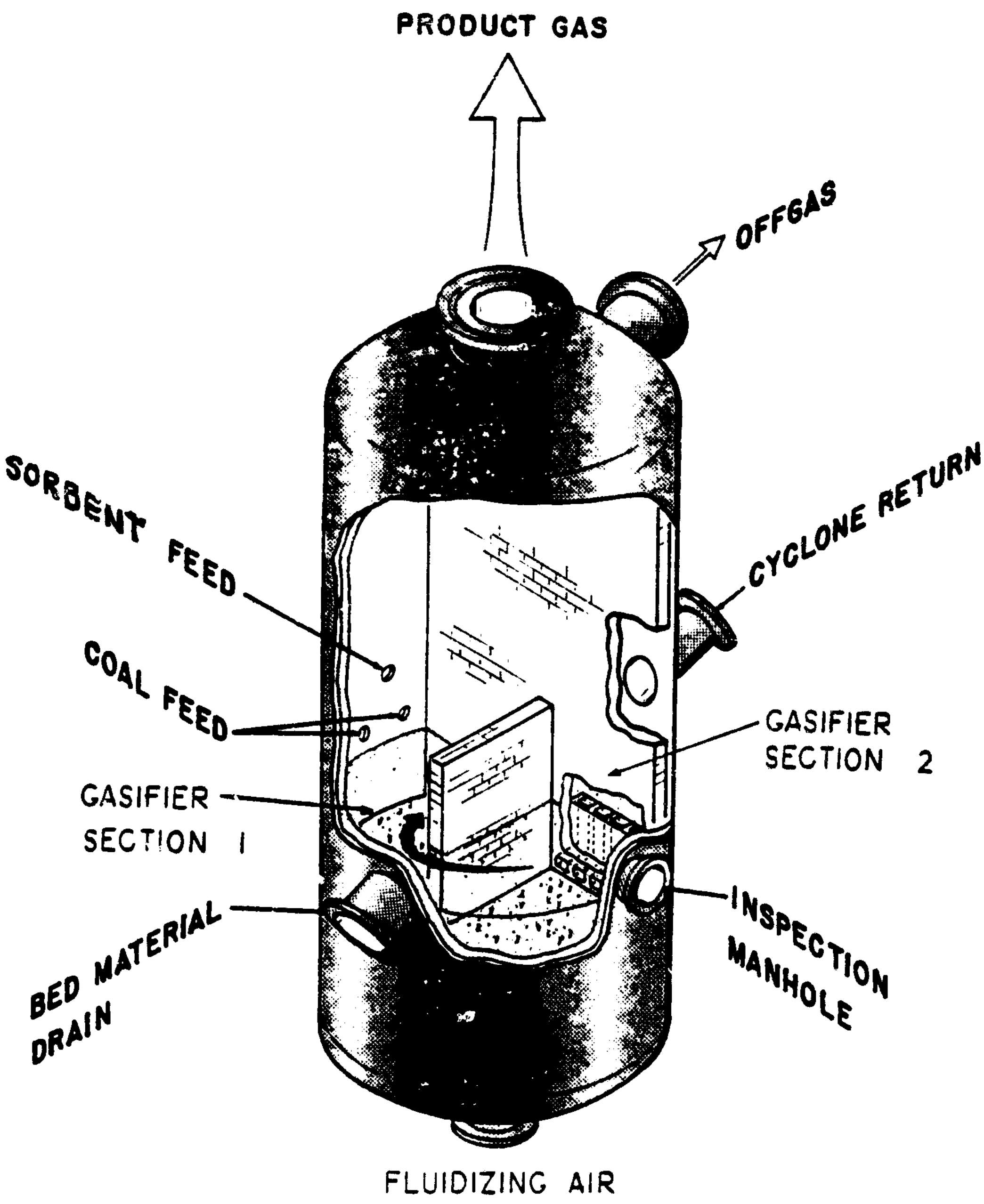


Figure 3.1-8. Foster Wheeler Chemically Active Pressurized Fluidized Bed

and drying requirements, etc.). Because of this complexity it is highly desirable to be able to scale combustor/gasified heat losses from a reference design operating condition as a function of mass flow rate, operating pressure and operating temperature so as to allow direct optimization of the net power output from the MHD flow train without repeating the combustor analysis. Heat loss scaling factors were derived from the following basic types of combustion/gasification systems:

1. Gas phase or suspension type reactors where the vessel length-to-diameter ratio (L/D) and gas residence time (τ) are fixed during scaling.

2. Vortex or cyclone type reactors where the vessel length-to-diameter (L/D) and characteristics gas velocity (u) are fixed during scaling.

3. Fixed bed, slowly moving bed or fluidized bed reactors where the bed height (L) and superficial gas residence time (τ) are fixed during scaling.

The above scaling criteria are compatible with present combustor/gasifier design practice and led to the development of heat loss scaling factors for fuel loading ($\dot{m}_f$), operating pressure (P) and operation temperature (T) which are shown in Table 3.1-5. In this table the subscript (W) refers to conditions at the walls. The variables labeled (1) refer to reference design conditions and the variables labeled (2) refer to operating conditions perturbed from the reference design. Combustion product mass flow ($\dot{m}_p$) is assumed proportional to the fuel input rate ($\dot{m}_f$), and the perturbations in mass flow, operating pressure and operating temperature are assumed not to produce major changes in combustion/gasification performance (e.g., turbulent mixing, kinetics, etc.) or in characteristic heat transfer coefficients.

Summaries of the reference design conditions for these three MHD combustor concepts are shown in Table 3.1-6. The reference heat loss and pressure drop data for the single-stage vortex combustor and the two-stage cyclone combustor have been developed from ETF scale design data presented in References 3 and 15, respectively. The data presented for the moving bed gasifier have been extrapolated from information provided in References 6 and 7.

For all three combustor concepts, scale-up to the commercial size plant is assumed to be by the addition of ETF size modular units with the first stage vessel sizes remaining fixed. Hot manifold heat losses are not accounted for in the single-stage vortex combustor concept since data on the vessel arrangement and/or vessel sizes for a commercial size plant are not available. Interstage manifold heat losses are included as part of the second stage heat loss for both the two-stage cyclone concept and the S^3PMB concept. For all concepts, the reference stagnation pressure and temperature at the exit of the MHD combustion system are 7.0 atm and 2800 K, respectively. To allow for wall heat transfer scaling, reference wall temperatures have been selected to reflect expected operating conditions for a given type of combustor. Second-stage wall temperatures are set at 1900 K while first stage wall temperatures are adjusted downward to correspond to lower operating chamber temperatures.

Table 3.1-5. Summary of Heat Loss Scaling Factors for Various Combustors/Gasifiers

Scaling and Type Combustor	Scaling Variable	Constraint Parameters	Heat Loss Scaling Factor
I <u>Fuel Loading Scaling</u>			
A. Gas Phase Combustor	$\dot{m}_f$	P, T, L/D, τ	$[\dot{m}_p(1)/\dot{m}_p(2)]^{2/3}$
B. Surface Burning or Cyclone Type Combustor	$\dot{m}_f$	P, T, L/D, U	constant
C. Moving Bed or Fluidized Bed Combustor	$\dot{m}_f$	P, T, L, τ	$\left[\frac{\dot{m}_p(1)}{\dot{m}_p(2)}\right]^{1/2}$
II <u>Pressure Scaling</u>			
A. Gas Phase Combustor	P	$\dot{m}_p$, T, L/D, τ	$[P(1)/P(2)]^{2/3}$
B. Surface Burning or Cyclone Type Combustor	P	$\dot{m}_p$, T, L/D, U	$[P(1)/P(2)]$
C. Moving Bed or Fluidized Bed	P	$\dot{m}_p$, T, L, τ	$[P(1)/P(2)]^{1/2}$
III <u>Temperature Scaling</u>			
A. Gas Phase Combustor	T	$\dot{m}$, P, L/D, τ	$\left[\frac{T(2)}{T(1)}\right]^{2/3} \left[\frac{T(2) - T_w(2)}{T(1) - T_w(1)}\right]$
B. Surface Burning or Cyclone Type Combustor	T	$\dot{m}$, P, L/D, U	$\left[\frac{T(2)}{T(1)}\right] \left[\frac{T(2) - T_w(2)}{T(1) - T_w(1)}\right]$
C. Moving Bed or Fluidized Bed Combustor	T	$\dot{m}$, P, L, τ	$\left[\frac{T(2)}{T(1)}\right]^{1/2} \left[\frac{T(2) - T_w(2)}{T(1) - T_w(1)}\right]$

Symbol Key: $\dot{m}_f$ = fuel input mass flow rate (fuel loading)

$\dot{m}_p$ = combustion products mass flow rate

P = operating chamber pressure of reactor vessel

T = operating chamber temperature of reactor vessel

T_w = reactor vessel wall temperature

L = characteristic reactor vessel length or bed height

D = characteristic vessel diameter

U = characteristic gas velocity in reactor vessel

τ = gas plug flow residence time or superficial gas residence time

(1) = variable value at reference design conditions

(2) = variable value at new or perturbed design point

Table 3.1-6. Reference Design Criteria for MHD Combustors and Nozzle Heat Loss Scaling

CONCEPT	STAGE	TYPE COMBUSTOR/GASIFIER	% HL	% P	% SLAG REJECTION	P atm	T_{gas} °K	T_{wall} °K	$\dot{m}_{gas}$ Kg/S	REMARKS
SINGLE-STAGE VORTEX	1st Stage	Single-Stage Vortex	5	5	70	7.0	2800	1900	438.5	%HL = f($\dot{m}$) = constant, modular unit scaling, vessel size fixed.
	2nd Stage	N/A	-	-	-	-	-	-	-	No accounting for manifold heat losses.
TWO-STAGE CYCLONE	1st Stage	Clustered Cyclones	5	8	85	7.6	2020	1800	-	%HL = f($\dot{m}$) = constant, modular unit scaling, vessel size fixed.
	2nd Stage	Gas Phase Combustor	2	2	-	7.0	2800	1900	438.5	Manifold heat losses included with 2nd stage.
SPLIT-STREAM SLAGGING MOVING BED	1st Stage	S^3PMB/SPMB	3.5	8	99.5	7.6	2110	1700	-	Modular unit scaling, vessel size fixed.
	2nd Stage	Gas Phase Combustor	2	2	-	7.0	2800	1900	438.5	Manifold heat losses included with 2nd stage.
	Nozzle	N/A	2.3	-	-	7.0	2800	1900	438.5	Scaling same as gas phase combustor.

I $$\%HL = 5.0\left(\frac{P(1)}{P(2)}\right)\left(\frac{T(2)}{T(1)}\right)\left(\frac{T(2) - T_w(2)}{T(1) - T_w(1)}\right) + 2.3\left(\frac{\dot{m}(1)}{\dot{m}(2)}\right)^{1/3}\left(\frac{P(1)}{P(2)}\right)^{2/3}\left(\frac{T(2)}{T(1)}\right)^{2/3}\left(\frac{T(2) - T_w(2)}{T(1) - T_w(1)}\right)$$

II $$\%HL = 5.0\left(\frac{P(1)}{P(2)}\right)\left(\frac{\tilde{T}(2)}{T(1)}\right)\left(\frac{\tilde{T}(2) - \tilde{T}_w(2)}{T(1) - T_w(1)}\right) + (2.0 + 2.3)\left(\frac{\dot{m}(1)}{\dot{m}(2)}\right)^{1/3}\left(\frac{P(1)}{P(2)}\right)^{2/3}\left(\frac{T(2)}{T(1)}\right)^{2/3}\left(\frac{T(2) - T_w(2)}{T(1) - T_w(1)}\right)$$

III $$\%HL = 3.5\left(\frac{\tilde{\dot{m}}(1)}{\tilde{\dot{m}}(2)}\right)^{1/2}\left(\frac{\tilde{P}(1)}{\tilde{P}(2)}\right)^{1/2}\left(\frac{\tilde{T}(2)}{\tilde{T}(1)}\right)^{1/2}\left(\frac{\tilde{T}(2) - \tilde{T}_w(2)}{\tilde{T}(1) - \tilde{T}_w(1)}\right) + (2.0 + 2.3)\left(\frac{\dot{m}(1)}{\dot{m}(2)}\right)^{1/3}\left(\frac{P(1)}{P(2)}\right)^{2/3}\left(\frac{T(2)}{T91)}\right)^{2/3}\left(\frac{T(2) - T_w(2)}{T(1) - T_w(1)}\right)$$

NOTE: ~ indicates first stage conditions.

3.1.4 RESULTS OF SPECIAL COMBUSTION STUDIES

The results and conclusions presented in this section pertain primarily to the HTAH and MHD combustors. However, it is necessary to emphasize that because of the close coupling between the HTAH, the MHD combustor, the MHD generator, and seed recovery and/or fuel gas clean-up subsystems a comparison of the performance of individual components will not necessarily correlate with changes and/or differences in overall system efficiency. The coupling effects between the MHD combustor and MHD generator are particularly strong, and therefore, results and conclusions presented in this section will somewhat overlap and should be put into context with results and conclusions in Section 3.3.

One important general observation is that heat loss scaling as a result of changes in plant size, operating pressure and plasma temperature significantly impacts overall system performance and costs. Another observation is that the composition of the plasma exiting the MHD combustor plays an important role with regard to overall plant efficiency and plant operation.

Both observations may appear to be somewhat obvious but, because of subtle interactions between combustor design and heat loss, operating conditions, plant size, plasma properties and MHD generator performance, they merit further discussion.

3.1.4.1 Comparative Heat Losses in 1 and 2 Stage Combustors

Typical results showing the variation in combustor and nozzle heat losses as a result of simultaneous changes in operating pressure and chamber temperature are given in Figure 3.1-9. Data are shown both for a single-stage vortex combustor with 70% slag rejection* (dashed line) and a two-stage cyclone combustor with 85% slag rejection (solid line).

The nozzle and second stage heat losses, which scale in proportion to $(P(1)/P(2))^{2/3}$, are less sensitive to pressure changes than either first stage or single-stage heat losses which scale in proportion to $(P(1)/P(2))$. For all combustors considered, increased pressure tends to increase the plasma temperature, but increased stagnation temperature tends to also increase heat loss, which in turn tends to restrain any temperature increase. Thus, there is a compensating effect between changes in heat loss and chamber temperature. The first stage heat loss for the two-stage combustor shown is less than the single stage combustor heat loss at the reference pressure of 7.0 atm because temperature scaling was not included for the first stage of the two-stage combustor. (It was assumed that, to control slag vaporization, the first stage temperatures would be maintained at a fixed level by slight adjustments to the fuel-to-air ratio). Slight changes in mass flow as a result of adjustments to the fuel-to-air ratio will not significantly impact the first stage heat loss as this scaling factor is essentially constant. At a given pressure, the total heat loss for the single-stage combustor is only about 1% less than the two-stage combustor and decreases with increasing plant size and increasing

* The reference design slag rejection for the single-stage vortex combustor is 70%. To permit an evaluation of the effects of heat loss for a given mass flow and slag rejection level, a case analysis was conducted assuming a slag rejection level of 85% to allow a comparison with the two-stage cyclone combustor concept.

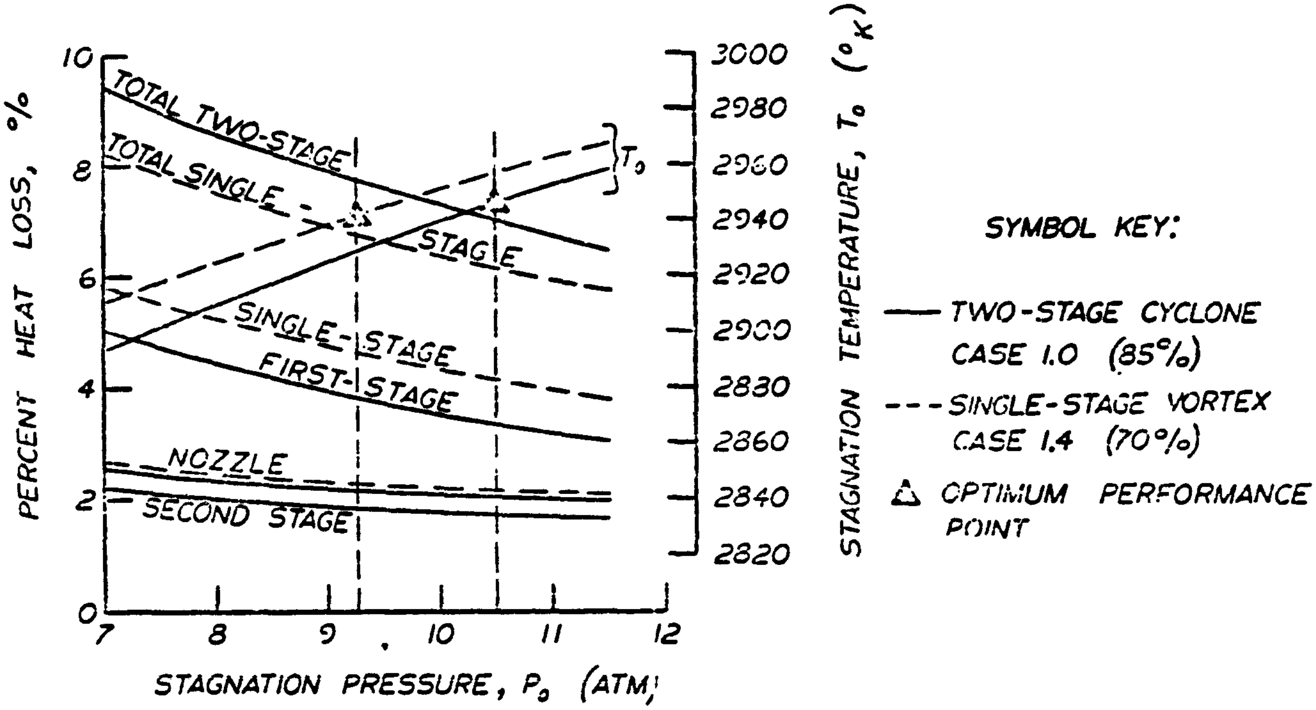


Figure 3.1-9. Effect of Pressure on Heat Loss for Single-Stage and Two-Stage MHD Combustors

pressure. Furthermore, for the single-stage concept these estimates do not include manifold losses which would arise if a modular design were necessary.

3.1.4.2 Considerations of Slag Rejection and Plasma Composition

The level of slag rejection indirectly affects combustor heat loss and hence combustor performance via alterations in operating pressure resulting from the optimization of MHD generator performance. For example, in comparing the combustor performance for Case 1.0 with that for Case 1.4, for optimized MHD generator performance, a two-stage cyclone combustor with 85% slag rejection has a slightly higher final stagnation temperature (T_o = 2945 K) than does a single-stage vortex combustor with 70% slag rejection (T_o = 2939 K). This result stems primarily from the fact that the operating pressure for optimum generator performance is shifted by the level of slag rejection. The optimum stagnation pressure for the single-stage combustor in this case is 9.5 atm, while the optimum pressure for the two-stage combustor is 10.4 atm. The increased operating pressure for the two-stage combustor does have a tendency to reduce the plasma conductivity, but the level of slag rejection has a more dominant impact on conductivity than does the pressure effect. For Case 1.0 the conductivity is 9.6 mho/m, while for Case 1.4 the conductivity is 9.8 mho/m.

The gross effects of slag rejection on conductivity can be seen in Figure 3.1-10 which presents conductivity data for the combustion products of Montana Rosebud as a function of plasma temperature. At 2800 K and 5 atm pressure, equilibrium analysis indicates the combustion of Montana Rosebud with total slag carryover yields a plasma with a conductivity of 12.2 mho/m. With 99.5% slag rejection, the conductivity is increased to about 15 mho/m.

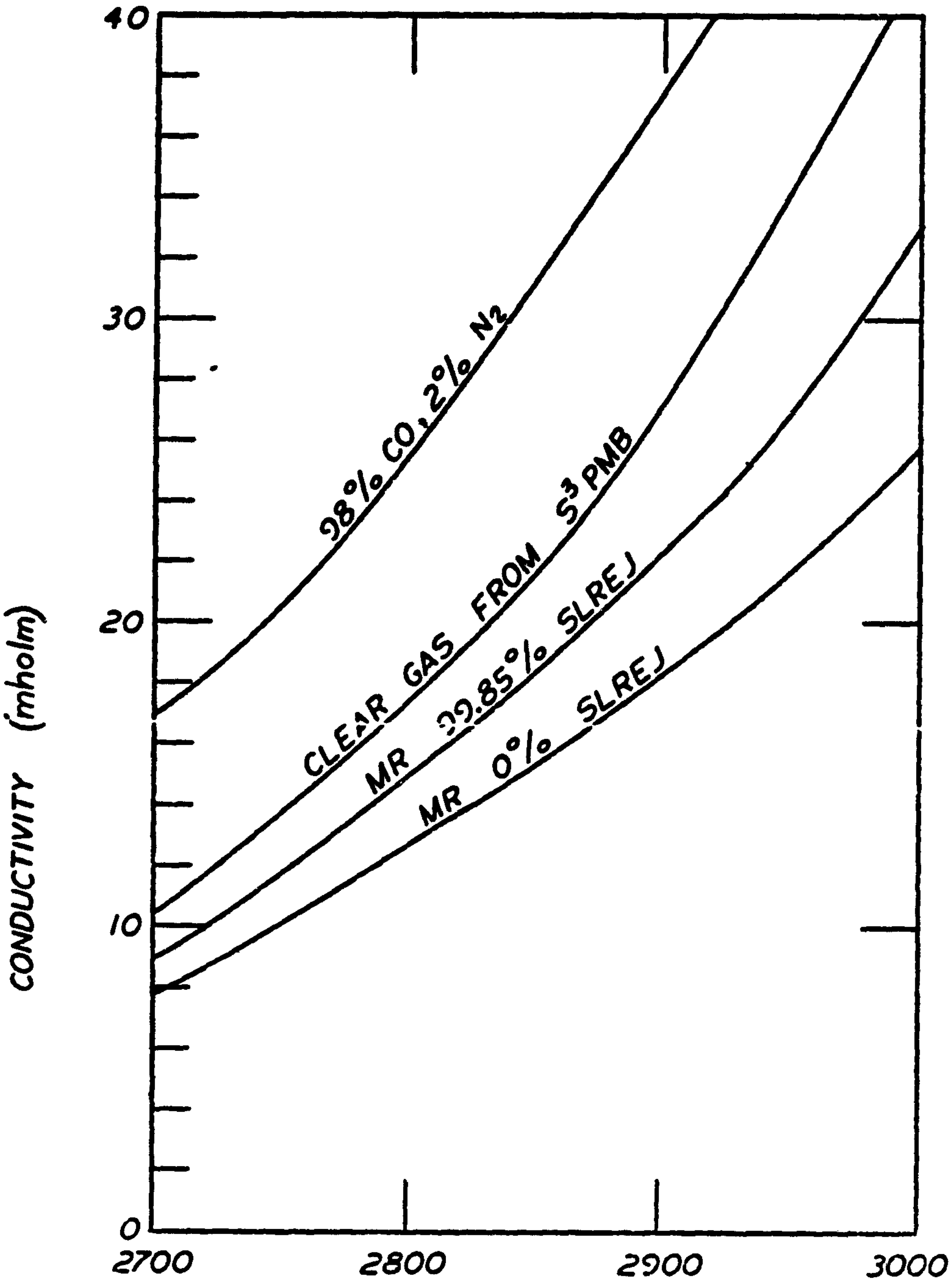


Figure 3.1-10. Combustion Product Conductivity for Several Fuels
E = 1.11, P = 5.0 atm, Temperature (°K)

3.1.4.3 Effect of S^3PMB 'Clear Gas' on MHD Generator Performance

Because of the adverse effects of fuel moisture, slag and other fuel components which degrade MHD generator output, the possibility of using CO rich 'clear gas' from a split-stream, slagging pressurized moving bed gasifier (S^3PMB) to enhance MHD generator performance was investigated (see Section 3.1.2.4). The 'clear gas' has good conductivity properties as the fuel for the MHD generator, and the 'top gas' which is high in hydrocarbons and contains released fuel moisture is suitable for the fuel for firing the HTAH. Conductivity data for a clear gas from an S^3PMB is given in Figure 3.1-10. The calculated MHD generator performance using this clear gas is excellent, but it is difficult to obtain an appropriate energy match between the top gas and clear gas to satisfy both the HTAH and MHD Combustor. Energy demands for the HTAH relative to the MHD combustors are on the order of 1:2; whereas, the energy split between the top gas and clear gas is on the order of 1:1. The energy split between top gas and clear gas, therefore, results in excess thermal energy delivered to the HTAH.

There are several options available to accommodate this energy mismatch. One approach is to find a use for the excess thermal energy delivered to the HTAH. In this case possibilities might include the firing of a separate conventional steam plant or use of the hydrocarbon-rich fuel as a feed stock for a chemical process plant. Another approach is to blend the 'clear gas' from the S^3PMB with the 'top gas' from a SPMB or other supplemental fuel so that a proper energy match for the HTAH and MHD combustor exists. In this latter case, the MHD generator performance is degraded by blending of the fuels. The analysis for Cases 2.0(a) and 2.0(b) indicate that the use of a clear gas alone results in about 19% more net power out from the MHD generator than the combined use of a S^3PMB and a SPMB. In addition, the data indicate that the direct use of pulverized coal (with total slag carryover) as a supplemental fuel for the S^3PMB is clearly not desirable since the net electrical output is about 20% less than the combined use of a S^3PMB and SPMB. As a point of reference, the S^3PMB/SPMB approach has about 37% less net MHD power out than the use of a single-stage vortex combustor concept (Case 2.0). The results of this investigation, therefore, indicate that the S^3PMB (as presently configured) does not offer a significant performance improvement over the use of single-stage vortex or two-stage cyclone combustors. It should be noted, however, that the present S^3PMB concept limited gasification air temperatures to 1300°F (978 K). If this air preheat temperature can be raised (e.g., to 2700°F), the performance of the S^3PMB may be substantially better than the use of single-stage or two-stage cyclone type combustors. Personal communications with the Grand Forks Energy Technology Center indicate that the operation of a slagging moving bed gasifier under these conditions may be possible. Further investigation, however, would be required to verify this mode of operation.

3.1.4.4 Effect of Coal Composition on MHD Generator Performance

Coal composition can significantly impact net MHD generator electrical output. This is due primarily to variations in inherent moisture content, heating value, slag content, and hydrogen content of the coal. High levels of fuel moisture and/or hydrocarbons which yield high concentrations of OH^- are not desirable because of the poisoning effect of this negative ion on plasma conductivity. Fuels with high heating value are generally desirable because of their high flame temperature capability, but consideration must also be given to slag content, slag

composition, and fuel composition. A comparison of Case 1.0 with 1.1 indicates that about 6% additional net electrical power is delivered by the MHD generator if Illinois #6 is used versus Montana Rosebud.

The effect of coal type on MHD performance is sufficient enough to encourage an examination of other coal types in addition to Montana Rosebud and Illinois #6. Preliminary analysis indicates that substantial improvements may result from use of higher rank coals. For example, compare the conductivity data for anthracite combustion with the conductivity data for Montana Rosebud and Illinois #6 as shown in Figure 3.1-11. In addition to a higher conductivity at a given temperature, higher flame temperatures can be reached with anthracite.

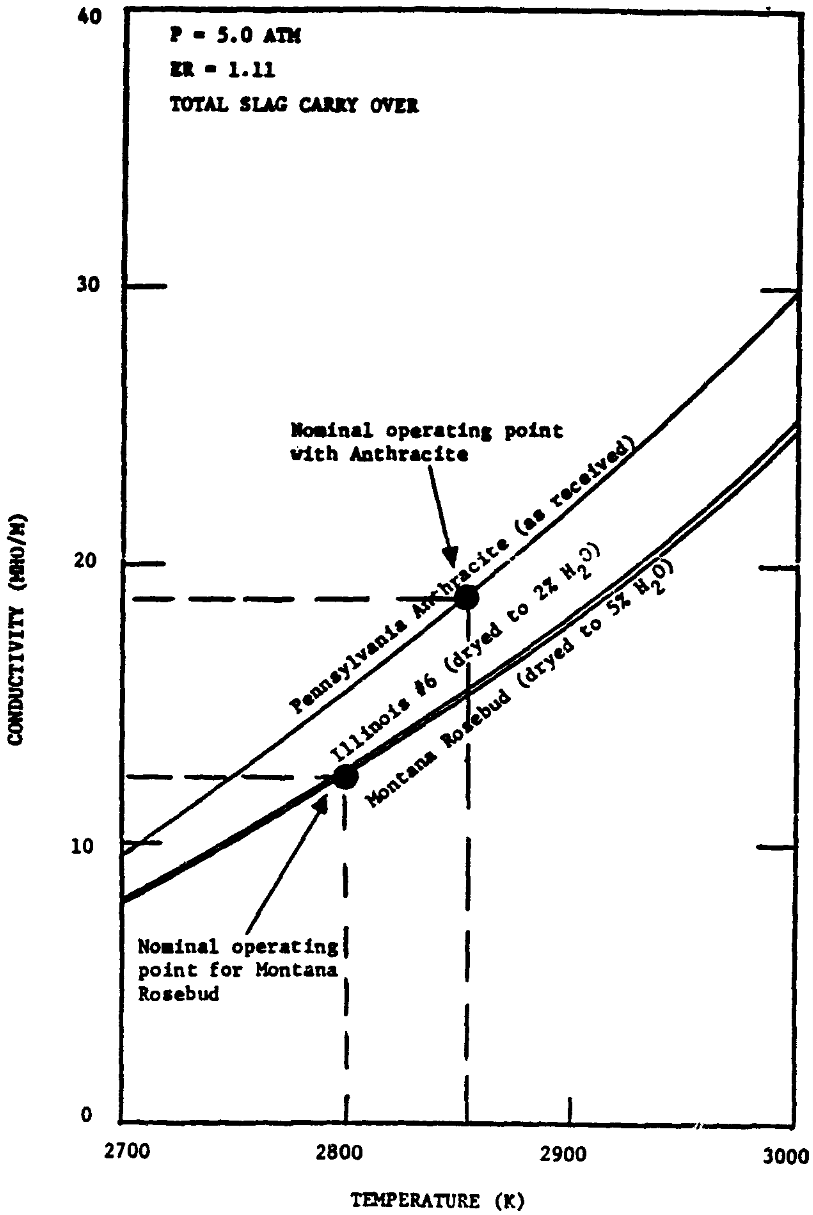


Figure 3.1-11. Effect of Coal Type on Plasma Conductivity

3.2 HIGH TEMPERATURE AIR HEATERS AND OXYGEN PRODUCTION

3.2.1 SYSTEM ASPECTS

The overall system efficiency of an MHD/steam plant with indirectly-fired high temperature air heaters (HTAH) is degraded from the directly-fired HTAH case because the former configuration involves less internal energy regeneration than the latter. To minimize the indirectly-fired penalty requires efficient utilization of the energy delivered to the preheat combustors, both in and downstream of, the HTAH. In addition to energy utilization considerations, the design of the preheat combustor flow train must provide for environmental control, e.g., NO_x, sulfur, and particulates.

The primary focus in minimizing the efficiency reduction of the indirectly-fired system is to minimize fuel requirements for the preheat combustor. The following arrangements were evaluated for this purpose:

1. Minimizing the energy that must be added to the MHD combustor air in the HTAH by maximizing the inlet air temperature. This configuration was examined in Case 2.16, where the inlet air temperature to the HTAH was raised to 1300°F by first heating the air in a (metallic) heat exchanger in the MHD flow train. The 1300°F air inlet case is termed the "hot-bottom" HTAH.

2. Displacing fuel energy from the preheat combustor with sensible heat of the air and recirculated flue gas fed to the preheat combustor. This configuration was examined in Case 2.18, where the temperature of the preheat combustor air and recirculated flue gas was elevated to 1300°F in metallic heat exchangers located in the MHD flow train.

Both arrangements described above involve regeneration of thermal energy from the MHD flow train downstream of the channel in the HRSR subsystem.

The second consideration affecting efficiency of the HTAH system is energy recovery from the combustion gas after it exits from the HTAH. Typical gas exit temperatures from the HTAH vary from 1500°F (hot bottom HTAH) to ~ 900°F. This energy must be recovered, down to a gas temperature of ~ 300°F, in downstream components.

Both atmospheric and pressurized combustion systems were examined for the HTAH system, the former in all of Base Case 1 and Case 2.12, and the latter in all of the other Base Case 2 configurations. Examples of both systems are shown in Figures 3.2-1 and 3.2-2.

For the pressurized arrangement, Figure 3.2-1, the gas inlet pressure is set equal to the air exit pressure, e.g., both the blowdown and reheat fluids are at the same pressure at the top of the ceramic matrix. The exit pressure of the air is determined by the pressure drop through the MHD components (combustor, nozzle, channel and diffuser), which then establishes the pressure of the reheat flow train. The combustion system is pressurized via the preheat compressor (PHC) which supplies the oxidizer to the preheat combustor (CB1, CB2). A high slag rejection, 2-stage, combustion system was used for all preheat combustors (except Cases 2.12

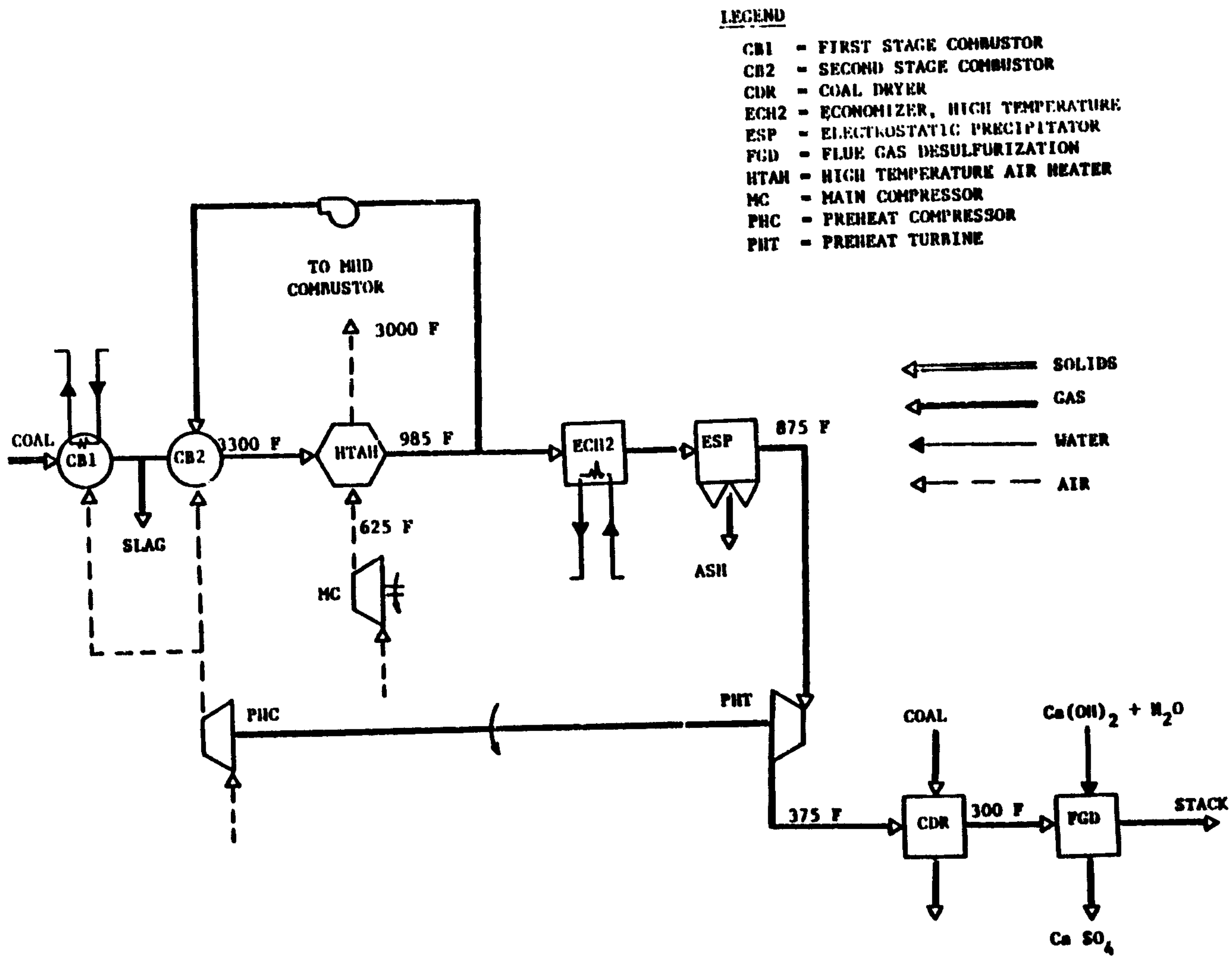


Figure 3.2-1. High Temperature Air Heater System, Case 2.0

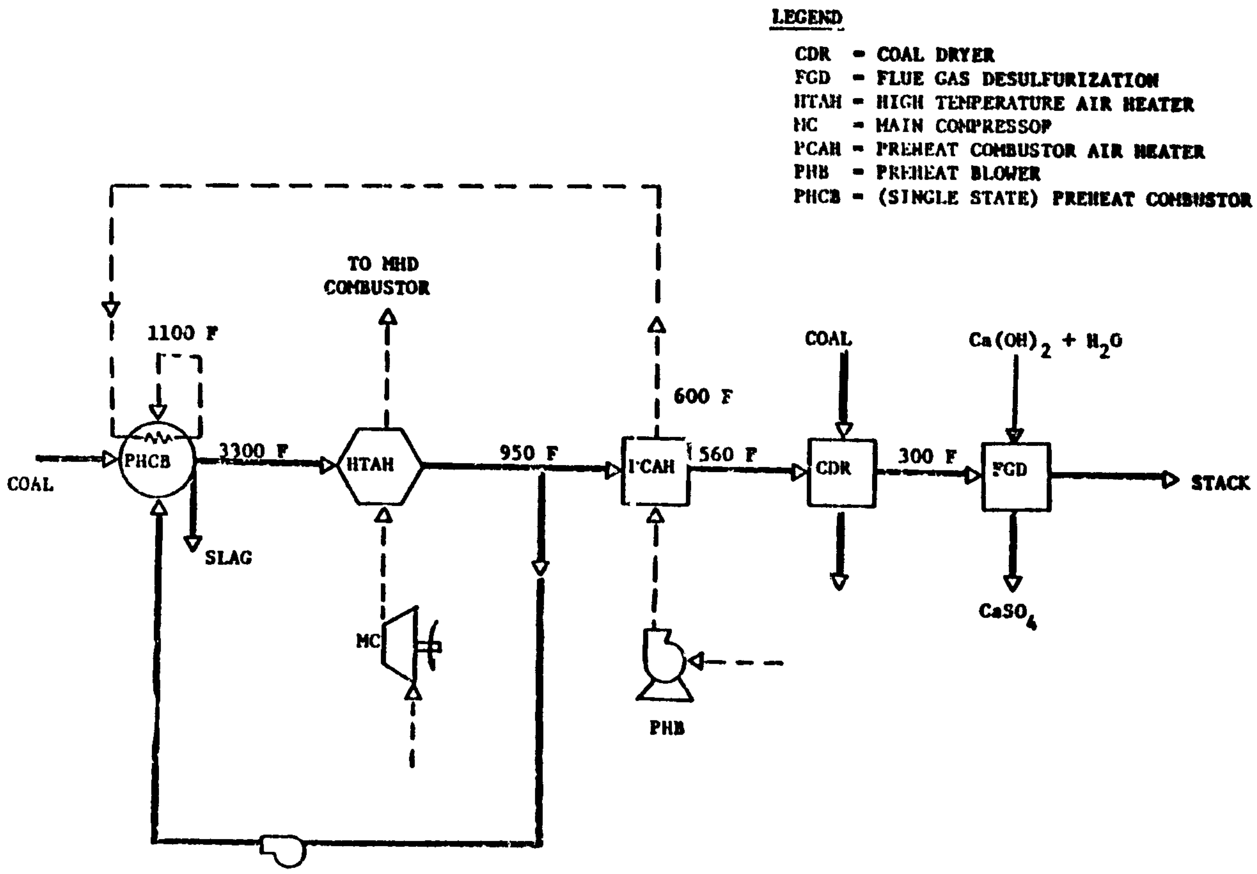


Figure 3.2-2. High Temperature Air Heater System, Case 2.12 (1 atm Combustor)

and 2.18) to minimize fouling of the HTAH ceramic matrix with slag. The combustion gas flow from the bottoming of the HTAH is split, part being recirculated to the combustor for NO_x reduction and the remainder entering the downstream components.

The arrangement of downstream components is also shown in Figure 3.2-1 for Case 2.0. Other modes of heat recovery between the HTAH exit and ESP inlet were used in some of the other cases. A modest amount of energy is first extracted from the gas in an economizer (ECH2) to adjust the inlet temperature to the gas turbine (PHT), such that the turbine output power matches the input power requirements of the preheat compressor. This arrangement is termed a balanced compressor/turbine set. The flow, before entering the turbine, is cleaned of the remaining ash in an electrostatic precipitator. After pressure let-down and energy extraction in the turbines, the low-grade gas energy is used for coal drying (CDR), with the gas entering the flue gas desulfurization equipment, a spray dryer (described in Section 3.8), at ~ 300°F.

The atmospheric combustion system for the HTAH system is illustrated, with an example, in Figure 3.2-2. Here, two types of combustors were evaluated. In all of the Base Case 1 design cases, a two-stage combustion system was used which incorporates a Wellman-Galusha gasifier. In Case 2.12, a single-stage, regeneratively air-cooled combustor was used*. This combustor is presently under development by General Electric specifically for the purpose of coal-firing a ceramic, regenerative heat exchanger, and has demonstrated ~ 90% slag rejection. A two-stage cyclone combustor, used in the majority of Base Case 2 pressurized HTAH systems, was found to be inappropriate for Case 2.12 because of excessive heat losses at atmospheric pressure. Downstream heat recovery is via an air preheater for the preheat combustor and coal dryers, with flue gas desulfurization before exhaust to the stack. The spray dryer system, because it incorporates gas clean-up equipment, also removes any remaining ash from the stack gas.

3.2.2 AIR HEATER CONFIGURATION

3.2.2.1 "State-of-the-Art" 600°F to 2700°F HTAH Subsystem with Atmospheric Pressure Reheat

The SOA HTAH assemblies were extensions of concept designs developed by the Arthur McKee Company as a subcontractor to GE in the ETF Study[1]. The delivered air temperature was 2700°F with inlet air temperatures of about 600°F. For the ETF, the high temperature air heater size and performance characteristics were based on results of McKee's Hot Blast Stove Computer Program, which has been verified by extensive field tests on existing hot blast stove installations. Operating and design data for the ETF SOA HTAH are given in Appendix 17 of Reference 1, including a comparison between McKee design data and world wide iron and steel hot blast stove design practice.

The 23' O.D. HTAH vessels of the ETF Study were scaled to 34' O.D. consistent with current hot blast stove designs. Internal firing with low Btu gas replaced external firing with fuel oil. Adjustments for the use of the low Btu gas were done using General Electric's Regenerative

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* A pressurized version of this combustor was used for Case 2.18.

Heat Exchanger (RECUP) and CCE Computer Codes. Checks on analysis of heat exchangers with the same design criteria indicate reasonable agreement between the GE and McKee codes.

Cases 1.0 and 1.1 were reference designs for Base Case 1 with other cases treated as perturbations. Reheat was at 1.15 atm and blowdown at 8.0 atm. Nominal air capacity of each vessel was 32.7 Kg/sec. Changes in blowdown pressure did not affect heat exchanger design since the pressure drop during reheat governs the heat exchanger design to a large extent. Pressure drop was limited to 15%, and heat losses to ambient were set at a nominal 5% of the total thermal input to the HTAH assembly. In order to limit the heat losses to 5% the combustion chambers must be integr... with the HTAH vessels and the steel pressure shells covered with insulation. Flue gas recirculation was used to control NO_x and to maintain the gas temperature at the top of the matrix at 3000°F (1922 K).

3.2.2.2 <u>600°F to 3000°F Air HTAH Subsystem with Atmospheric Pressure Reheat</u>

This HTAH concept is an extension of the SOA HTAH to higher temperature operation through the use of high quality (99.5% alumina versus 90% alumina) refractory materials. Capital cost considerations generally do not permit the use of advanced refractories with SOA heat exchanger designs. This concept was investigated to assess the effects of capital cost investment for higher quality refractories in a SOA heat exchanger and the corresponding impact of performance change in MHD power output as a result of higher preheat temperatures. A 3000°F exit air temperature is assumed, and the corresponding inlet air temperature is about 600°F. The pressure drop was limited to 4%, and the heat losses were set at 5% of the total thermal input for all operating pressures, which varied between 6.8 and 11.1 atm.

The analysis for the advanced air heaters in Base Case 2 was also conducted by application of the General Electric RECUP and CCE computer programs. For all advanced cases, flue gas recirculation was utilized to control NO_x and to maintain the gas temperature at the top of the HTAH matrix at 3300°F.

3.2.2.3 <u>600°F to 3000°F Air HTAH Subsystem with Pressurized Reheat</u>

This HTAH subsystem is an extension of the advanced pressurized reheat/pressurized blowdown concept developed during the GE ETF Study. As with the atmospheric pressure concept, the use of high density alumina (99.5% alumina) allows a maximum delivered air temperature of 3000°F. Some importance differences between this concept and those discussed above are as follows:

1. For the advanced ETF concept, high purity alumina refractory was used for the entire matrix column as well as for the core liner.

2. The advanced heat exchanger with pressurized reheat is designed for thermal stress limits rather than pressure drop limits. The use of high purity (99.5%) alumina refractories for both the matrix column and the hot core liner allows higher operating temperatures than a HTAH utilizing lower purity (90%) alumina. In addition, the thermal stresses are higher than for a HTAH with similar refractories but operating with atmospheric reheat. In essence, pressurized operation during both reheat and

blowdown allows a closer approach to a thermal stress limit design. Therefore, a more compact HTAH vessel can be designed. A theoretical elastic stress limit of 2000 psi was used in the design analysis for this advanced concept and is a level at which short term thermal stress cracking has not been observed with Norton AH 299A alumina cored brick in actual regenerative heat exchanger operation at GE.

3. Pressurized operation during reheat and blowdown has the added advantage of requiring almost no pressure differential across the hot blast valves which then serve only a stream isolation function. Matching the reheat and blowdown pressures also eliminates repressurization purge cycles.

3.2.2.4 1300°F to 3000°F Air HTAH Subsystem with Pressurized Reheat

Raising the temperature of the bottom of the heater to a temperature consistent with 1300°F inlet air permits greater thermal energy regeneration, hence higher plant efficiency. Also, the cost of the HTAH itself can be reduced by shortening of the bed height due to the decrease in air ΔT across the HTAH; i.e., (3000°F - 1300°F) versus (3000°F - 600°F). Recent work at the University of Montana and at FluiDyne Engineering Co. also indicates the possibility of eliminating the problem of flyash plugging by periodically heating the bottom of the unit to a temperature sufficient to melt out slag.

Detailed engineering designs for a hot bottom heat exchanger to be fired by a cyclone combustor with greater than 90% slag rejection are currently being developed by GE with design and consulting support from FluiDyne Engineering Co. The design incorporates a ceramic arch to avoid the necessity of a water cooled support grate.

3.2.3 OXYGEN PRODUCTION

The viability of using oxygen enrichment as a supplement to or replacement of the HTAH subsystem is highly dependent on the shaft power required for oxygen production. Since the objective is simply to increase O_2/N_2 ratio, production of pure O_2 is not essential and recent data from Lotepro, Inc., New York, N.Y. as furnished by NASA LeRC[2] indicates that O_2 gas can be produced for 197.8 KW-hr/ton of equivalent pure O_2 delivered at atmospheric pressure and ambient temperature. This is a substantial improvement over the value of approximately 300 KW-hr/ton used in the GE ETF study[1].

NASA LeRC also furnished cost data for the plant indicating a turnkey cost of $35 million for a 2000 Ton/day plant or $100 million for three such plants.

These data were used to incorporate O_2 production without attempting to integrate the O_2 plant into the rest of the facility.

Thermochemical O_2 production was not investigated but might provide a more efficient means of providing supplementary oxygen. Membrane diffusion is another technique for increasing the O_2/N_2 ratio but cost increases linearly with volume required and membranes are not competitive even at ETF size[3].

3.3 MHD GENERATOR

3.3.1 INTRODUCTION

Results are presented of calculations that determine the performance of the MHD generator component consistent with overall MHD/steam power plant system requirements. Special emphasis is given to the treatment of the coupling between the MHD generator and the combustor. This coupling occurs, for example, as a result of the dependence of the generator performance on combustor heat loss and the variation of the heat loss with the operating pressure, which, in turn, is determined from an optimization of the net MHD generator electrical power. The net MHD output electrical power is determined by subtracting from the gross MHD electrical output, the compressor power required and also, if applicable, the power to produce oxygen for oxygen enriched combustion.

The generator performance calculations to be discussed here are listed in Table 3.3-1; these cases cover the major parametric effects addressed in the study. The nominal conditions for the parametric study are listed in the footnote to Table 3.3-1, and the parametric effects studied in each case are explained in the far right-hand column. The use of O_2-enriched combustion is considered for Base Cases 1 and 3, and the common effects considered in each of the base cases are single-stage and two-stage cyclone combustors, Montana Rosebud (MR) and Illinois #6 (I6) coal, and increased magnetic field. In Base Case 2, the effects of Cs seed (Case 2.5), a supersonic channel (Case 2.6) and reduced thermal size (Cases 2.10 and 2.11) are studied. The effects of a S^3PMB gasifier are considered in Cases 2.0(a) and 2.0(b). Also in Base Case 2, the effects of a constant electrical stress channel are studied in Case 2.16 by allowing the magnetic field to vary as required to satisfy the constraint of a constant, maximum transverse electric field of $E_{y,max} = 4$ kV/m.

3.3.2 ANALYTICAL MODELS

The three essential analytical elements required to calculate generator performance are: (1) characterization of the thermodynamic and transport properties of the coal/air/seed combustion product working fluid; (2) determination of the combustor size, flame temperature and heat loss; (3) prediction of the electrical and gas dynamic performance of the MHD generator. The combustor analysis and the heat loss scaling relationship are discussed in Section 3.1. The analytical models used in this work for items (1) and (3) above are discussed in the following sections.

3.3.2.1 Coal/Air/Seed Working Fluid Properties

The coal/air/seed working fluid properties are calculated using the CCE (Coal Combustion Equilibrium) computer code, Appendix C.

For all of the cases considered, the overall fuel-to-oxidizer ratio (ϕ) for the coal combustion is determined by the relationship $\phi = \phi_S \phi_E$, where ϕ_S is the fuel-to-oxidizer ratio for stoichiometric combustion and ϕ_E is the equivalence ratio. A single value of the equivalence ratio is used in this study, which is $\phi_E = 1/0.9 = 1.1111$. This corresponds to a fuel rich condition involving 90% of the oxidizer required for stoichiometric combustion. For the cases

Table 3.3-1. Parametric Cases for MHD Generator/Combustor Performance Study

BASE CASE	PARAMETRIC CASE	COMBUSTOR/ GASIFIER CONCEPT	OVERALL SLAG REJECTION	PARAMETRIC EFFECT STUDIED
1	1.0	2-Stage Cyclone	85%	Reference Case
T_{AIR} = 2700 F	1.1	2-Stage Cyclone	85%	I6 Coal
	1.2	2-Stage Cyclone	85%	Nominal Air, R_{O_2} = 0
R_{O_2} = 10%	1.3	2-Stage Cyclone	85%	Increased Field, B = 7-6 Tesla
	1.4	1-Stage Vortex	70% and 85%	1-Stage Vortex
2	2.0	1-Stage Vortex	70% and 85%	Reference Case
	2.0a	S^3PMB + Coal	38%	0% Slag Rejection for Added Coal
	2.0a	S^3PMB + Coal	91%	85% Slag Rejection for Added Coal
T_{AIR} = 3000 F	2.0b	S^3PMB + SPMB	99.85%	Integrated Gasifiers for HTAH and MHD Flow Train
	2.1	1-Stage Vortex	70%	I6 Coal
R_{O_2} = 0	2.2	2-Stage Cyclone	85%	2-Stage Cyclone
	2.5	1-Stage Vortex	85%	Cs Seed
	2.6	1-Stage Vortex	85%	Supersonic Channel, M_1 = 1.2
	2.7	1-Stage Vortex	85%	Increased Field, B = 8-7 Tesla
	2.10	1-Stage Vortex	85%	Decreased Size, $P_{TH,C}$ = 1500 MW L = 15 m
	2.11	1-Stage Vortex	70% and 85%	Decreased Size, $P_{TH,C}$ = 2000 MW L = 20 m
	2.16	1-Stage Vortex	85%	Constant Electrical Stress, B Field Satisfies $E_{y,MAX}$ = 4 kV/m Constraint
3	3.0	1-Stage Vortex	70%	Reference Case
	3.1	1-Stage Vortex	70%	I6 Coal
T_{AIR} = 1300 F	3.2	2-Stage Cyclone	85%	2-Stage Cyclone
	3.4	1-Stage Vortex	70%	T_{AIR} = 1100 F
R_{O_2} = 40%	3.5	1-Stage Vortex	70%	Increased Field, B = 8-7 Tesla

*Unless noted otherwise, common conditions for all cases are: $P_{TH,C}$ = 2800 MW, MR Coal, Subsonic Channel (M_1 = 0.8, M_f = 1.0), L = 25 m, B = 6-5 Tesla, $C_{P,DIFF}$ = 0.6, V_{DROP} = 100 V

with O_2 enrichment, the percentage level of O_2 enrichment is specified by the value of the parameter R_{O_2}, defined as (# kg pure O_2/1 kg moist, nominal air) x 100.

The seed material is a mixture of K_2CO_3 and K_2SO_4 with just enough K_2CO_3 added to chemically combine with the elemental sulphur in the coal. The amount of seed material added is based on providing 1% K element by weight relative to the total coal/air/seed mixture for conditions of zero O_2 enrichment. This seed loading is approximately optimum for the coals and fuel-to-air ratio (90% of stoichiometric air) used here because larger seed fractions tend to reduce the combustor flame temperature and thus the electrical conductivity. The seed-to-coal mass ratio corresponding to the conditions of 1% K of zero O_2 enrichment is kept fixed for the cases with O_2 enrichment. This results in an element weight percentage greater than unity because, for the same thermal input to the combustor, the total coal/air/seed mass flow rate is lower than O_2-enriched air. Seed reprocessing results (Section 3.9) suggest the direct use of potassium formate as the seed material in future analyses.

3.3.2.2 Generator Performance Analysis

The MHD generator performance results have been obtained for a linear, Faraday channel using a quasi-one-dimensional, core flow plus integral boundary layer analysis. In its basic form, this analysis has been previously applied and validated in the generator performance study for ETF. The computer code utilized has since been modified to account for the combustor and generator coupling and to determine automatically a generator solution that satisfies certain prescribed constraint conditions.

Figure 3.3-1 shows schematically how the MHD generator analysis is coupled to the analyses describing the coal/air/seed working fluid properties and the combustion and nozzle heat loss scalling relationships. The combustor and nozzle are coupled to the generator through the influence of the pressure ($p_{o,i}$) on the combustor heat loss, which determines the delivered combustor temperature available to the nozzle and generator. The combustor heat loss also depends on the combustor temperature and the mass flow rate as well as on the pressure. Because the generator performance optimization discussed below involves essentially a search for an optimum operating pressure, it has been very advantageous in this parametric study to utilize the heat loss correlations developed here for the vortex, cyclone and gasifier combustors that account for the effects of pressure, temperature and mass flow rate.

For given coal/air/seed properties and the thermal input power to the combustor, ($P_{TH,C}$), the generator output obtained includes the gross electrical output power (P_{EL}), the operating stagnation pressure ($p_{o,i}$), the velocity gradient (u') and the generator area ratio (A_f/A_i) or the Faraday generator loading parameter (K). By an iteration of these latter three quantities, a generator solution is obtained that satisfies specified constraint values for exit stagnation pressure ($p_{o,f}$), maximum Hall field ($E_{x,max}$) and exit Mach number (M_f). This iteration is accomplished using a coupled Newton-Raphson solution approach with the necessary partial derivatives for the 3 x 3 solution matrix calculated numerically. Specifying the velocity gradient of the MHD core flow is advantageous because this removes the singularity in the differential equation that occurs at M = 1, and the numerical integration can be executed rapidly and without numerical difficulty for near-solic generator flow conditions. For best performance, the velocity should

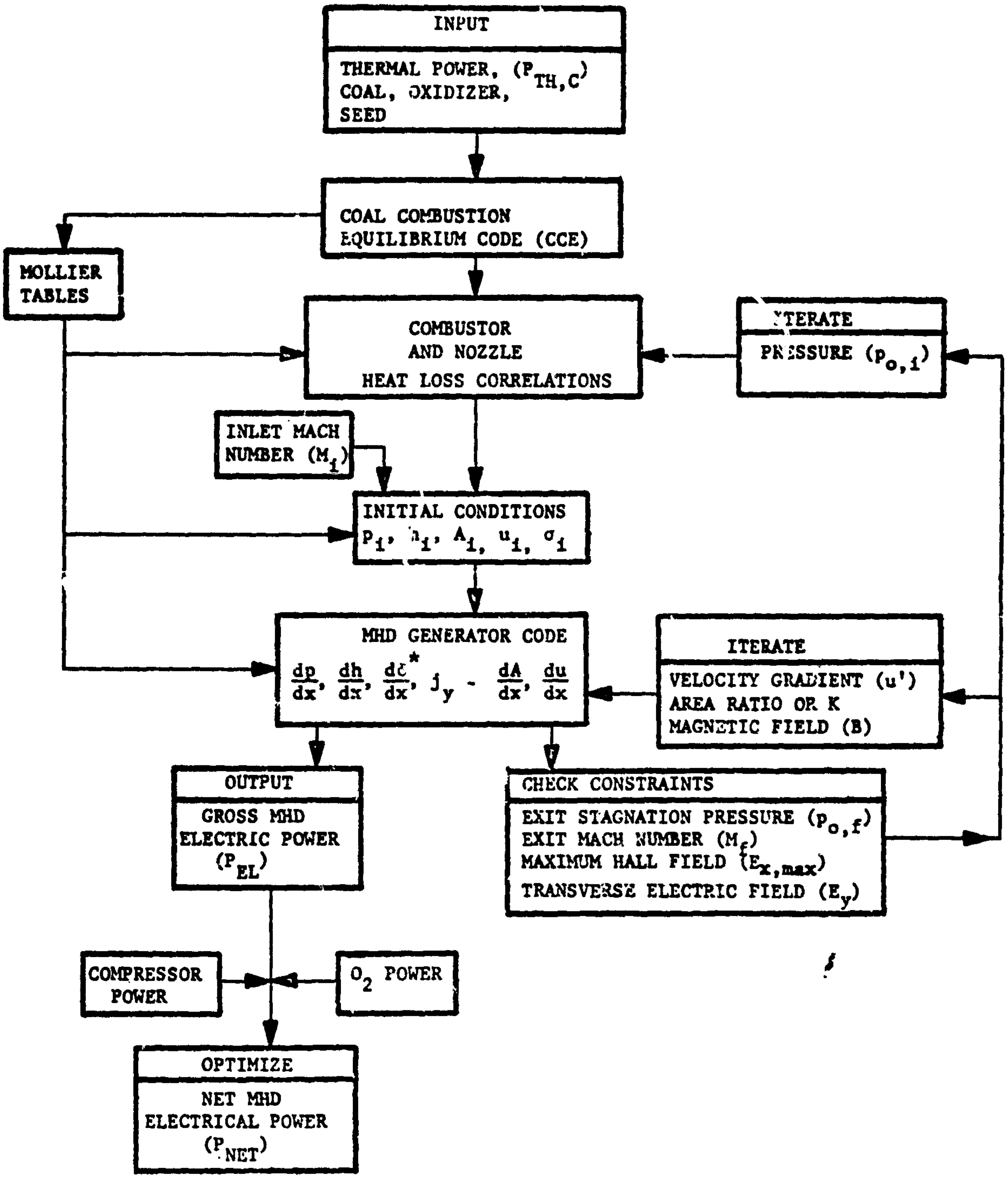


Figure 3.3-1. Schematic Approach for Coupled MHD Generator/Combustor Performance

be accelerating or decelerating for subsonic and supersonic generators, respectively. For both cases, the tendency is toward a sonic condition at the generator exit when electrical energy is extracted.

The analysis provides two options with regard to the channel area shape, A(x), and the Faraday loading parameter, K. One option allows the loading parameter to be input and the analysis computes the required channel area distribution. However, the nominal program option utilized here inputs the area shape and the analysis computes the loading distribution and the other solution parameters that are consistent with the specified constraint conditions. The area distribution utilized here is described by a third-order function that satisfies conditions on the entrance and exit area slopes along with values for the inlet area and the area ratio. Exit area is determined as part of the generator solution procedure. The selected channel shape, with small entrance and exit region slopes, gives near open circuit loading conditions and small axial field values in the end regions to minimize end losses associated with circulating Hall currents. Also, with the current density proportional to dA/dx a large channel exit slope can, in conjunction with the low exit region working fluid temperatures, cause the Hall field to increase to unacceptably high levels, leading to a compromise in the generator performance.

The nominal magnetic field distribution used for the performance calculations has a 6-5 Tesla, linearly-tapered, active field region bounded by entrance and exit fringe field regions of one meter length over which the field decreases linearly to a value of 2 Tesla. This nominal field distribution is shown in Figure 3.3-2 compared with the Case 2.16 magnetic field distribution, determined to satisfy a constraint of E_y = 4 kV/m. The Case 2.16 solution for the magnetic

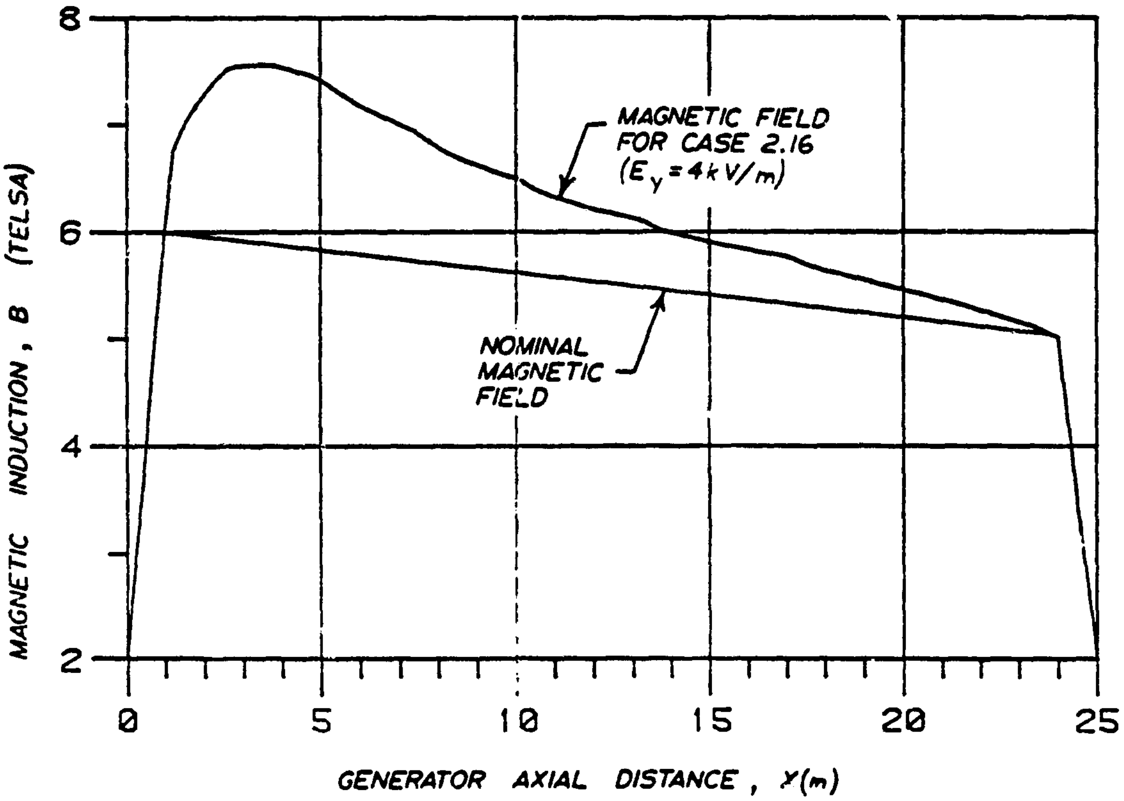


Figure 3.3-2. Comparison of Nominal and Calculated Magnetic Field Distribution

induction (B) is obtained from the Ohm's law relation $uB + j_y/\sigma = E_y = 4$ kV/m. This is actually a nonlinear equation in B because, in the present formulation with the velocity gradient specified, $j_y \sim 1/B$.

The performance results show that there is an operating point where the net MHD generator power is a maximum. Although the gross MHD electrical output power generally increases with operating pressure (for fixed exit pressure), the rate of this increase eventually becomes less than that of the required compressor power, and the net MHD electrical power passes through a maximum. For such a situation, the net generator power is said to be "compressor-power-limited." Because the Hall field also tends to increase with operating pressure, a maximum allowable Hall field value can occur before the maximum net power point is reached, and for this situation the net MHD power is said to be electric-stress-limited. As an example of compressor-power-limited performance, Figure 3.3-3 shows the calculated performancr results for Case 1.0. The performance optimization approach involves obtaining generator solutions for a range of operating conditions spanning either the compressor-power or the electric-stress-limited operating point. As shown in Figure 3.3-3, the optimum net electrical power is a relatively weak function of the operating pressure. Thus, it may be advantageous to accept a small performance penalty by operating at a pressure higher than the optimum in order to take advantage of potential lower combustor and air heater costs resulting from reduced physical sizes at higher pressure.

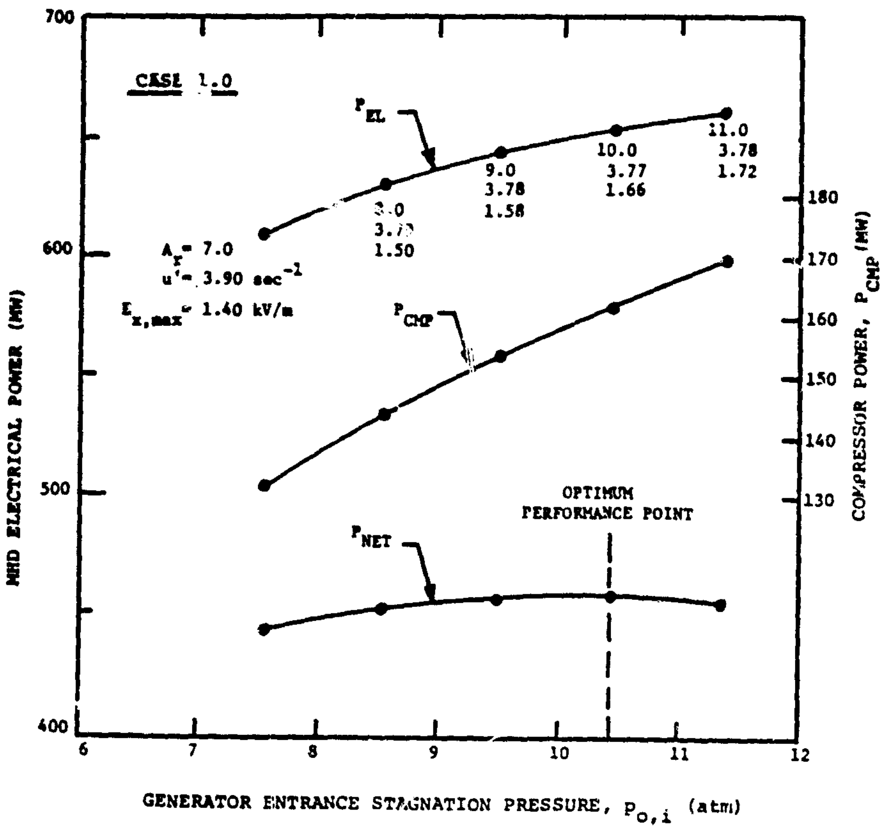


Figure 3.3-3. Case 1.0 Results Illustrating Optimization of Net MHD Electrical Power

For purposes of the generator performance optimization studies, the single-stage, specific compressor power is expressed in terms of the compressor pressure ratio and isentropic efficiency. The isentropic efficiency is calculated as a function of the pressure ratio and the polytropic efficiency, with the latter assigned a value here of 0.89. To account for pressure losses between the compressor and the generator, the compressor outlet pressure is taken to be ten percent higher than the inlet operating pressure of the MHD pressure ($p_{o,i}$). To calculate the required O_2 power for the O_2 enriched cases, a specific O_2 power of 0.786 MW-sec/kg is used.

3.3.3 PERFORMANCE RESULTS

The performance study results for selected parameters are presented in Tables 3.3-2, 3.3-3, and 3.3-4, Base Cases 1, 2 and 3, respectively. The values tabulated are the MHD generator gross electrical output power (P_{EL}), the net generator power (P_{NET}), the delivered combustor stagnation temperature (T_o), the generator entrance stagnation pressure ($p_{o,i}$), the generator entrance electrical conductivity (σ_i) and the maximum axial (Hall) and transverse electric field values ($E_{x,max}$ and $E_{y,max}$), respectively.

Table 3.3-2. Base Case 1 MHD Generator Performance

Case *	P_{EL} (MW)	P_{NET} (MW)	T_o (K)	$p_{o,i}$ (atm)	σ_i (mo/m)	$E_{x,max}$ (kV/m)	$E_{y,max}$ (kV/m)	ΔP_{NET} (%)
1.0	654.2	459.8	2945	10.44	9.56	1.66	4.14	Reference
1.1	691.6	489.2	3000	11.38	11.51	1.52	4.18	+6.39
1.2	590.3	430.4	2789	7.65	5.99	2.77	3.98	-6.39
1.3	702.7	499.6	2956	11.41	9.37	1.89	4.96	+8.66
1.4 (70%)	634.6	449.2	2939	9.48	8.76	1.72	4.14	-2.31
1.4	664.4	469.2	2955	10.44	10.04	1.60	4.15	+2.20

* All cases are for 85% slag rejection unless noted otherwise.
All cases are with air + 10% O_2 unless noted otherwise

3.3.3.1 Base Case 1

An O_2 enrichment level of R_{O_2} = 10% is used for all the Base Case 1 cases except for Case 1.2, which does not use O_2. This level is optimum value for the conditions of Case 1.0 as determined by calculations for Case 1.0 with varying O_2 enrichment. It has been assumed that a 10% O_2 enrichment level is approximately optimum for the other Base Case 1 cases as well. Table 3.3-2 shows that the range in the predicted gross electrical power for the Base Case 1 cases is from 634.6 MW to 702.7 MW, while the net power range is from 449.2 MW to 499.6 MW. The O_2 power is 32.7 MW for all the O_2 enrichment cases so that the compressor power for the combustion air accounts for most of the difference between the net and the gross power values.

Table 3.3-3. Base Case MHD Generator Performance

Case*	P_{EL} (MW)	P_{NET} (MW)	T_o (K)	$P_{o,i}$ (atm)	σ_i (mho/m)	$E_{x,max}$ (kV/m)	$E_{y,max}$ (kV/m)	ΔP_{NET} (%)
2.0 (70%)	623.0	469.5	2856	7.64	7.27	2.29	4.08	Reference
2.0	658.3	491.4	2874	8.64	8.25	2.19	4.09	+4.66
2.0a (38%)	477.6	366.5	2722	4.51	4.09	3.05	4.01	-21.96
2.0a (91%)	607.2	456.0	2780	6.67	7.40	2.66	3.98	-2.88
2.0b (99.85%)	603.6	448.7	2738	6.67	7.17	2.80	3.94	-4.43
2.1 (70%)	663.7	497.0	2905	8.65	8.45	2.14	4.13	+5.86
2.2	636.7	475.9	2854	8.14	7.79	2.27	4.07	+1.36
2.5	720.6	541.9	2886	9.66	11.27	1.70	4.09	+10.28**
2.6	605.1	471.9	2838	6.25	6.16	3.00	6.09	-3.97**
2.7	723.8	556.3	2875	8.70	8.24	3.00	5.71	+13.21**
2.10	298.6	223.4	2846	6.73	8.41	2.66	3.99	-54.54**
2.11 (70%)	410.5	311.4	2840	6.62	7.33	2.45	4.04	-33.67
2.11	437.4	327.5	2861	7.65	8.36	2.36	4.06	-33.35**
2.16	715.0	531.1	2891	10.13	7.92	2.85	4.07	+8.08**

*All cases are for 85% slag rejection unless noted otherwise.

**Reference is Case 2.0, P_{NET} = 491.4.

Table 3.3-4. Base Case 3 MHD Generator Performance

Case*	P_{EL} (MW)	P_{NET} (MW)	T_o (K)	$p_{o,i}$ (atm)	σ_i (mho/m)	$E_{x,max}$ (kV/m)	$E_{y,max}$ (kV/m)	ΔP_{NET} (%)
3.0	560.1	350.6	2946	9.30	8.55	1.37	4.12	Reference
3.1	613.5	386.3	3000	11.71	10.40	1.34	4.17	+10.18
3.2 (85%)	585.1	361.4	2964	11.61	9.24	1.39	4.13	+ 3.08
3.4	550.6	337.5	2925	9.30	7.65	1.47	4.11	- 3.74
3.5	658.3	421.9	2990	12.95	8.05	1.86	5.76	+20.34

* All cases are for 70% slag rejection unless noted otherwise
All cases with air + 40% O_2

As noted in Table 3.3-2, a 6.39% decrease in net electrical power is predicted for Case 1.2 relative to the Case 1.0 reference value of 459.8 MW. This decrease is the result of the lower delivered temperature (2789 K vs 2945 K) without O_2 enrichment and the lower resulting generator entrance conductivity. The Case 1.2 maximum Hall field value of 2.77 kV/m is largest among the Base Case 1 cases because Case 1.2 has the lowest working fluid temperatures. The largest parametric effect for Base Case 1 is the 8.66% increase in net power for Case 1.3, which has a 7-6 Tesla field instead of the nominal field with a 6-5 Tesla taper. This power increase is due, in part, to a larger transverse electric field ($E_{y,max}$ = 4.96 kV/m), which varies as the product uB. Also, the higher field shifts the point of optimum generator power to a higher pressure and this has the beneficial effect of reducing the combustor heat losses.

A relatively large net power increase of 6.39% is also predicted using I6 coal (Case 1.1). The high temperatures and conductivities for Case 1.1 are the combined result of a larger HHV and a lower moisture content for I6 coal compared to MR coal. As indicated in Tables 3.3-3 and 3.3-4, similar effects are predicted for the Base Case 2 and 3 cases considering I6 coal. Table 3.3-5 summarizes the predictions for Base Cases 1-3 to show the effect of coal type on net MHD electrical power. For all three base cases, using I6 coal gives an approximately 30 MW increase in net power; however, on a percentage basis, the net power increase is 10.18, 6.39 and 5.86 percent for Base Cases 3, 1, and 2 respectively. Since the corresponding O_2 enrichment levels are 40, 10 and 0 percent, this result appears to suggest that a larger relative effect of I6 coal can be expected when O_2 enrichment is employed.

The Base Case 1 results indicate a slightly lower performance for a single-stage vortex combustor with a 70% slag rejection (Case 1.4) than for a two-stage cyclone combustor with 85% slag rejection (Case 1.0). The coupled heat loss scaling and slag rejection phenomena are responsible for the predicted results and important considerations for the design of the MHD generator and combustor components. Similar trends are noted in the Base Case 2 and 3 results that compare combustor type and slag rejection, and a separate discussion of these effects is given in a section following the discussions of the other Base Case 2 and 3 results.

Table 3.3-5. Effect of Coal Type on Net MHD Electrical Power, P_{NET} (MW)

CASE	MR COAL			I6 COAL		
BASE CASE	(CASE)	70%*	85%**	(CASE)	70%*	85%**
1	(1.0)	---	459.8	(1.1)	---	489.2
2	(2.0)	469.5	---	(2.1)	497.0	---
3	(3.0)	350.6	---	(3.1)	396.3	---

* Single-Stage Vortex
** Two-Stage Cyclone

3.3.3.2 Base Case 2

The Base Case 2 parametric cases which consider the effects of combustor type (Cases 2.0, 2.0(a), 2.0(b), 2.2) and coal type (Case 2.1) are discussed elsewhere. This section discusses the other Base Case 2 cases, which consider the effects of cesium seed (Case 2.5), supersonic generator (Case 2.6), reduced thermal size (Cases 2.10 and 2.11) and magnetic field (Cases 2.7 and 2.16). Each of these cases utilizes a single-stage vortex combustor with 85% slag rejection; therefore, as can be noted in Table 3.3-3, the indicated relative net power performance is based on using as a reference the Case 2.0, single-stage, 85% slag rejection prediction of 491.4 MW.

A relatively large 10.25% increase in net power is predicted for Case 2.5 with cesium seed. This is the result of a larger conductivity, which is indicated in Table 3.3-3 by the generator entrance value of σ_i = 11.27 mho/m. In addition, the larger pressure (9.66 atm) reduces the combustor heat loss and thus increases the delivered combustor stagnation temperature (2886 K). To obtain the cesium seed results, the calculated potassium element (K) conductivities are scaled, keeping the condition of 1% by weight of seed element and accounting for the effect of the differences in the Cs and K ionization potentials on the electron density.

The net power prediction for the supersonic channel (Case 2.6) is 471.9 MW, which is approximately 20 MW lower than the subsonic channel reference case. This 3.97% decrease in power is the result of lower generator temperatures caused by the initial supersonic expansion. The lower temperatures increase the Hall field and this requires operation at a lower area ratio and a lower pressure (6.16 atm) in order to limit the maximum Hall field value. The lower operating pressure tends to increase the combustor heat loss and this compounds the low temperature problem. Along with the Hall field, the maximum transverse electric field (6.09 kV/m) is highest for the supersonic channel case. The 471.9 MW net power value is the electric-stress-limited value. Approximately 12 MW more power could be achieved by operating on a higher pressure at the compressor-power-limited optimum point, but the corresponding maximum Hall field value would then be about 3.6 kV/m.

The largest effect among the Base Case 2 cases with $P_{TH,C}$ = 2800 MW is the 13.21% net power increase predicted for Case 2.7, which has an 8-7 Tesla magnetic field. The larger field increases the induced transverse electric field and this increases the output power. The 3 kV/m maximum Hall field value indicates an electric-stress-limited performance; however, the electric-stress-limited operating point and the compressor-power-limited optimum power point are nearly identical so that the potential net power increase is only a few megawatts larger than the 556.3 MW value noted. Cases 1.3, 2.16 and 3.5 also consider the effects of magnetic field. Unlike the Case 2.7 result, the net power increases predicted for Cases 1.3 and 3.5 are not electric-stress-limited. This is because the generator temperatures are higher for the Base Case 1 and Base Case 3 conditions and this keeps the Hall field low. The approximately 72 MW net power increase noted for Case 3.5 is the largest observed among the cases compared. The Case 2.16 performance corresponds to a magnetic field determined to satisfy a constraint of a constant transverse electric field. The predicted net power is 513.1 MW, which is a performance increase of 8.08 % compared to the reference case with a 6-5 Tesla field. This is a very attractive case considering that excessive electric stress conditions are not required to achieve this performance, the maximum Hall field and transverse field values being 2.85 kV/m and 4.07 kV/m, respectively. The magnetic field required to achieve the Case 2.16 performance is shown in Figure 3.3-2. The maximum field value is 7.56 Tesla, occurring at an axial distance of 3.2 m.

The effects of thermal size are investigated in Base Case 2 for Cases 2.10 and 2.11, which correspond to thermal inputs and generator lengths of 1500 MW, L = 15m and of 2000 MW, L = 20m, respectively. The shorter channel lengths are a compromise between obtaining a high enthalpy extraction and having an excessive length to diameter ratio for a given thermal size. Relative to the 491.4 MW net power value for the 2800 MW reference case, the performance decreases are 33.35% and 54.54%, respectively, for Cases 2.11 and 2.10. This variation is approximately linear so that the net MHD power can be expressed as a linear function of the thermal input power over the range from 1500 - 2800 MW with a maximum error of only 2 - 3%.

3.3.3 BASE CASE 3

The performance results for Base Case 3 are all obtained for a 40% O_2 enrichment. This is an optimum value as determined in calculations for case 3.0, and it is assumed to very nearly optimum for the other Base Case 3 cases. As shown in Table 3.3-4, the net power predictions for Base Case 3 are about 100-120 MW less than the results for comparable cases in Base Cases 1 and 2, but the corresponding differences in the gross electrical power predictions are smaller. The required O_2 power for Base Case 3 of about 85 MW is responsible for the reduced net electrical power. The delivered combustor stagnation temperatures and the operating pressures are generally highest for Base Case 3. The high temperatures keep the conductivity high and the Hall field low, and the high pressures are advantageous with regard to combustor heat losses that determine the delivered stagnation temperature. The effects represented by Cases 3.0, 3.1, 3.2 and 3.5 are discussed elsewhere. Case 3.4 considers the effect of a decrease in the air preheat temperature value from 1300K to 1100K. The predicted effect is a 3.74% decrease in the net power. Although smaller net power is predicted for Base Case 3, the overall system efficiency can still be comparable to that for Base Cases 1 and 2 because of the extra fuel needed by the latter to satisfy the energy requirements of the indirectly-fired high temperature air heater components.

3.3.4 COMBUSTOR TYPE AND SLAG REJECTION EFFECTS

Table 3.3-6 summarizes the results obtained for Base Cases 1, 2 and 3 considering the effects of combustor type and slag rejection. Single-stage vortex and two-stage cyclone combustors are compared in all three base cases, and Base Case 2 considers in addition a S^3PMB gasifier/combustor. For all three base cases the predicted performance for the single-stage vortex and the two-stage cyclone combustor is comparable; however, the interesting result is obtained that the single-stage vortex performance with 70% slag rejection is less than that of the two-stage cyclone with 85% slag rejection. The gasifier performance is the lowest of the three combustor types considered, but the performance for the best gasifier case is only 2.88% below that of a single-stage vortex with 70% slag rejection. This result is obtained for Case 2.0(a) (91%) and is perhaps optimistic because of the higher 85% slag rejection assumed for the added coal. The Case 2.0(a) gasifier results are seen to be very sensitive to the overall slag rejection rate.

Table 3.3-6. Effect of Combustor Type and Slag Rejection on Net MHD Electric Power, P_{NET} (MW)

CASE	SINGLE-STAGE VORTEX			TWO-STAGE CYCLONE		S^3PMB GASIFIER			
BASE CASE	(CASE)	70%	85%	(CASE)	85%	(CASE)	99.85%	91%	38%
1	(1.4)	449.2	469.9	(1.0)	459.8	---			
2	(2.0)	469.5	491.4	(2.2)	475.9	(2.0a)	--	--	366.5
						(2.0a)	--	456.0	--
						(2.0b)	448.7	--	--
3	(3.0)	350.6	--	(3.2)	361.4	--			

The slag rejection effects are also evident in the single-stage vortex combustor comparisons for Base Cases 1 and 2. For Base Case 1, the effect on the net power of increasing the slag rejection from 70% to 85% is 4.5% (Case 1.4), and is 4.66% (Case 2.0) and 5.17% (Case 2.11) for Base Case 2. This slag rejection increase for the single-stage vortex combustor is sufficient to boost its performance above that of the two-stage cyclone. The slag rejection effect on performance is the result of the adverse effect on electrical conductivity of ash species negative ions, which are present in larger concentrations as the slag rejection is decreased. The magnitude of the effect is such that in spite of lower heat loss, the net power predictions for a single-stage vortex combustor with 70% slag rejection are lower than those for a two-stage cyclone with 85% slag rejection. The combustor and generator coupling is important because the reduced conductivity lowers the value of the pressure giving optimum generator performance and this tends to increase the combustor heat loss, which scales inversely with the pressure.

Comparing the single-stage (70%) and two-stage (85%) net power predictions given in Table 3.3-5, it is seen that there is approximately a ten (10) MW advantage for the two-stage combustor for each of the three base cases. However, the corresponding percentage increases are 1.36, 2.36 and 3.08 percent for Base Cases 2, 1 and 3, respectively, indicating an increased relative effect of combustor types for cases with O_2 enrichment.

The effects of slag rejection and the generator/combustor coupling are dramatically demonstrated by comparing the Case 2.0(a) results for 38% and 91% overall slag rejection shown in Table 3.3-6. The net power prediction for a 38% slag rejection is smaller by almost 90 MW. This large performance reduction is caused, in part, by the decreased conductivity associated with a low slag rejection. The lower conductivity tends to increase the Hall field and this reduces the performance by requiring a lower channel area ratio and lower operating pressure to limit the Hall field. The lower pressure, in turn, increases the gasifier combustor heat loss, which scales inversely with pressure, and this compounds the original conductivity effect by lowering the delivered working fluid temperature. It is noted that the performance for this case is electric-stress-limited; that is, higher net MHD powers are possible at higher pressures, but this is not allowed because the corresponding maximum Hall fields are greater than the imposed design limit of 3 kV/m.

3.3.5 THERMAL EFFICIENCY SENSITIVITY

To estimate the sensitivity of the thermal efficiency to variations in the net MHD power, a simplified energy balance expression for the thermal efficiency is expressed solely in terms of the net MHD power. To do this, appropriate values are assigned to the steam plant and the HTAH efficiencies and to a parameter accounting for cycle heat losses such as coal drying and stack losses. The gross electrical power predictions for a given base case are also correlated as a linear function of the net MHD electrical power. The resulting thermal efficiency expression is a linear function of the net power, and this expression is differentiated to give the desired sensitivity. Expressed as the percent change in thermal efficiency per unit percent change in the net MHD power, the sensitivity results obtained for Base Cases 1, 2 and 3 are 0.069, 0.066 and 0.058, respectively. Thus, a 10 percent change in net electrical power results in a 0.6 - 0.7 percent change in the overall thermal efficiency. Since the variation in the net electrical power is in the 5 - 10 percent range for the majority of the parametric cases considered in this study, the corresponding overall differences in thermal efficiency fall within the 0.5 - 1.0 percent range. Thus, the thermal efficiency is seen to be relatively insensitive to the MHD generator performance for the magnitude of the parametric effects predicted in this study. Among the cases considered, however, the best performance should be obtained for the Base Case 2 system concept with I6 coal, a two-stage cyclone combustor with 85% slag rejection, a subsonic generator, and a magnetic field distribution consistent with a constant transverse electric field of 4 kV/m.

3.3.6 GENERATOR AND COMBUSTOR PERFORMANCE CONCLUSIONS

The major conclusions of the generator and combustor performance study are as follows:

1. For the majority of the cases considered, the optimum net MHD generator power is compressor-power-limited rather than electric-stress-limited.

2. The optimum net power is a relatively weak function of the operating pressure and it may be advantageous from the overall systems perspective to operate on the high pressure side of the optimum point to take advantage of potential lower costs of the combustor and air heater components as a result of reduced sizes at high pressures.

3. The generator and combustor performance analyses have been coupled through the use of combustor heat loss correlations that scale the effects of pressure, temperature and mass flow rate. The dominant coupling mechanism is the result of a combustor heat loss that varies inversely with the pressure and the influence of this heat loss and of the combustor slag rejection on the value of the optimum generator operating pressure.

4. Comparable generator performance is obtained using single-stage vortex or two-stage cyclone combustors; however, the single-stage vortex performance is lower for a slag rejection of 70%, but it is higher for a slag rejection of 85%. The S^3PMV gasifier concept gives the lowest MHD generator performance among the combustor types considered. This is primarily the result of the gasifier's lower average air preheat temperature.

5. The influence of combustor slag rejection on MHD generator performance occurs as a result of an effect of ash species negative ions on electrical conductivity.

6. A 5 - 10 percent increase in net electrical power is predicted for Illinois #6 coal compared to Montana Rosebud coal. This indicates a relatively strong effect of coal type on MHD generator performance.

7. Generator performance increases on the order of 10 - 20% are predicted for an increase in the peak field from 6 Tesla to 8 Tesla. This performance increase occurs as a result of larger induced transverse electric fields which are in the 5 - 6 kV/m range.

8. For the nominal 2800 MW thermal size, the thermal efficiency is relatively insensitive to changes in the net MHD generator power. For the magnitude of the generator performance effects predicted here, the variation in thermal efficiency should not exceed 1 percent.

9. Among the system concepts and parametric cases considered, the best performance should be obtained for the Base Case 2 plant concept with 3000°F air preheat, zero O_2 enrichment, I6 coal, two-stage cyclone combustor, subsonic generator of 25 m length and a magnetic field distribution determined to give a constant transverse electric field of 4 kV/m.

3.4 MAGNET

Primary effort on magnet evaluation was the determination of size, weight and cost for all cases. Scaling was based on design data for the AVCO BL6-1 base load magnet[1] (a circular saddle design). This design was used since it is a base load design relatively close in size to the PSPEC requirements and because it was the only base load magnet of this size range on which significant design data was available.

Four cases were evaluated in some detail, by GE's Energy Systems Products Department (ESPD), three of them for a maximum field of 6 Tesla at sizes corresponding to nominal plant output power levels of 1257 MWe, 887 MWe and 655 MWe and one for a maximum field of 8T at a plant output of 1293 MWe. An estimate was then made by GE-ESPD of the change in weight and cost for +20% to -40% changes in warm bore diameter at the MHD generator exit and generator length. These data were then used to calculate magnet costs for other specific parametric cases.

In addition to the cases discussed above, a qualitative evaluation was made of the magnet required to provide the field specified for the constant electrical stress case (Case 2.16). The judgement was that while the calculated peak field near the channel entrance was somewhat in excess of 7 Tesla, it was 5 Tesla near the exit where the bore is largest and mechanical containment problems are greatest so that the 6 Tesla tapering to 5 Tesla case was reasonably representative. For costing purposes, this magnet was assumed for Case 2.16.

3.4.1 SPECIFICATION OF WARM BORE SIZE

In the absence of a channel design specifically for PSPEC, several previous studies[2,3,4] over a substantial range of thermal power and mass flow rate were utilized to plot a correlation of the ratio of warm bore size to channel width at MHD generator exit. Data were plotted as a function of mass flow rate, Figure 3.4-1, and an empirical equation used to calculate warm bore diameters.

$$R = 2.27 - 0.436\ (\log_{10}\dot{m} - 2.0) \qquad (1)$$

where R is the ratio of warm bore diameter to exit channel width and m is mass flow rate in Kg/sec. Based on this correlation, data shown in Table 3.4-1 were used for magnet evaluation. Nominal magnetic field distribution is shown in Figure 3.4-2.

3.4.2 SCALING AND COST ESTIMATION

Equations used for scaling and a Sample Case, are included as Appendix I. This section is limited to a brief general description.

The required ampere turns were scaled proportional to the product of on-axis field (B) and radius of windings (R). Ampere turn requirements were determined at entrance and exit of the channel and were scaled to meet the peak requirements at both ends. Ampere meters were scaled proportional to ampere turns and magnet length. Magnet build was scaled proportional to ampere turns and inversely to radius of winding. Mass and weight of conductor;

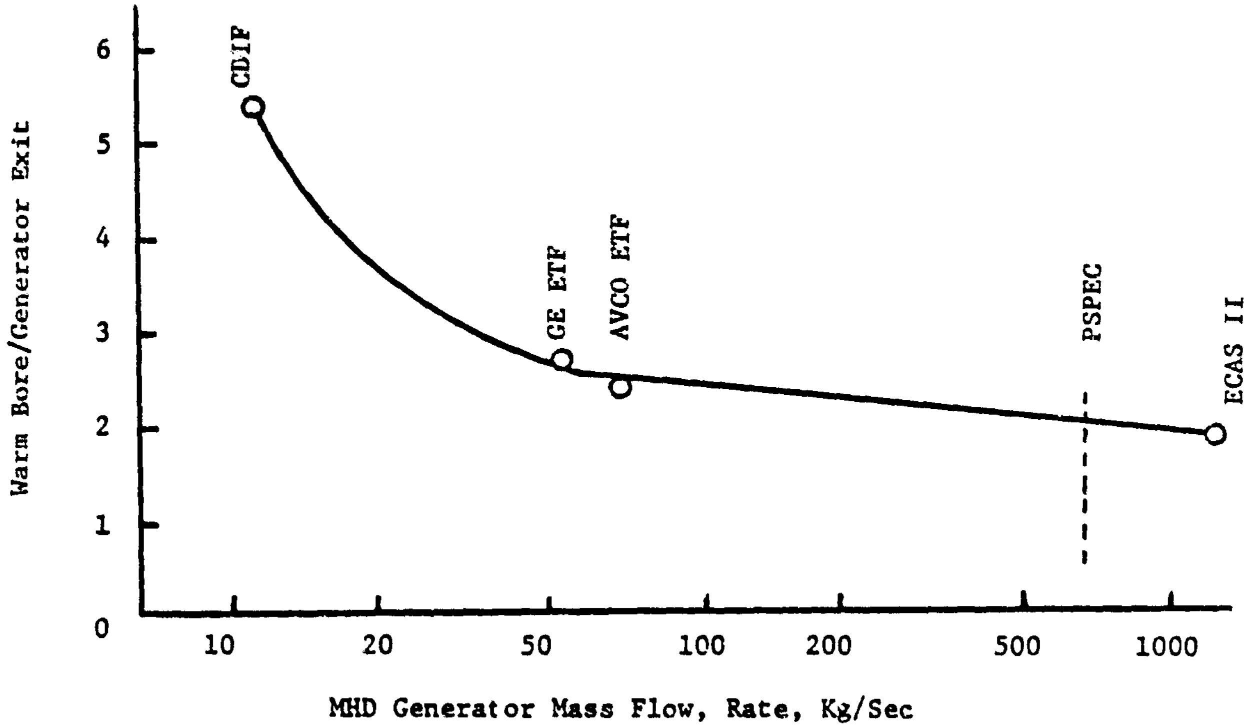


Figure 3.4-1. Correlation of Ratio Magnet Warm Bore to MHD Generator Exit with Mass Flow Rate

Table 3.4-1. Basic Size Data for PSPEC Magnets. Channels are Assumed Square, Magnet Bore Assumed Round

Case	Output Power MWe	Length m	B_1	B_{L-1}	Mass Flow Rate (Kg/sec)	Channel Entrance Width (m)	Channel Exit Width (m)	Warm Bore Diameter (m)
2.2	1257	25	6	5	606	1.04	2.8	5.4
2.11	887	20	6	5	434	0.90	2.4	4.8
2.10	655	15	6	5	325	0.83	2.0	4.1
2.7	1293	25	8	7	606	1.00	2.8	5.4

support shells, etc., were scaled proportional to build, winding radius and magnet length.

For the force containment structure the I-beam section modulus (Ik) was scaled proportional to the square of the product of magnetic field and radius of windings and the I-beam cross-section selected to provide needed section modulus. Structure mass and weight were scaled proportional to I-beam cross-section area, ring grider radius and magnet length. Thus, the mass of force containment structure is nearly proportional to total magnet energy stored (i.e , proportional to volume integral of square of magnetic field). The radiation shield, outside jacket, etc., were scaled as the product of magnet length and maximum diameter. The radiation shield, outside jacket, etc., were scaled as the product of magnet length and maximum diameter (i.e., scaled to outer surface area of magnet).

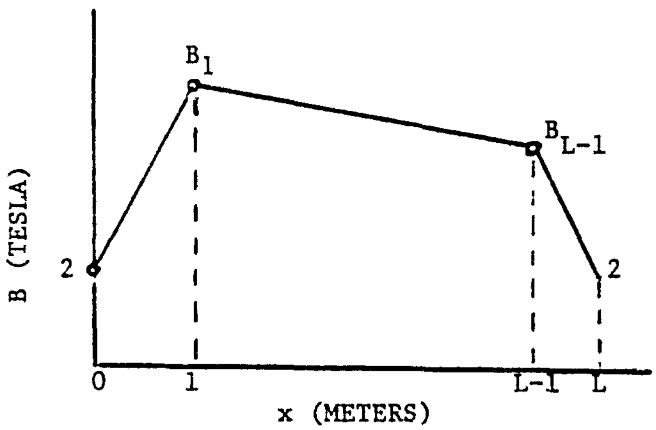


Figure 3.4-2. Nominal Field Distribution for Cases with Prescribed Magnetic Field

Costing was based on the following average component costs:

Conductor	\$20/kg to \$30/kg
Structure	\$10/kg
Cryostat	\$16/kg

Manpower costs were assigned by engineering judgement based on weight and dimensions of magnet.

Table 3.4-2 and 3.4-3 summarize component, weight and costs respectively for the four specific cases evaluated. Figure 3.4-3 and 3.4-4 show changes in magnet size, weight and cost for size perturbations about the 1257 MWe, 6T case. Sensitivity to changes in warm bore at the MHD generator entrance was also estimated but cost variation was less than 5% and judged negligible to the accuracy of the overall estimate.

3.5 POWER CONDITIONING AND INVERSION EQUIPMENT

From the Faraday channel calculations, a diagonal-connected-wall channel current distribtuion was selected which appeared likely to preserve tolerable Hall voltage gradients along the channel while, at the same time, producing current and voltage values at the power take-off taps which are conducive to an efficient and cost effective power conditioning system design. Figure 3.5-1 illustrates the selected DCW design*. Three current take-off taps centered around points 6, 10 and 14 meters from the channel entrance contribute about 8800,

*The specific design is based on MHD generator results for Case 2.2.

Table 3.4-2. Component Weight Summary (10^3 Kg)

	1287 MWe, 6T	887 MWe, 6T	655 MWe, 6T	1293 MWe, 8T
SUPERCONDUCTOR	28	22	15	148
STABILIZING COPPER	835	641	440	2171
SUBTOTAL, CONDUCTOR	863	663	455	2319
WINDING SUPPORT SHELLS	999	769	526	2597
GIRDERS	4508	3528	2548	11270
COIL CONTAINER	494	378	260	1284
INSULATION AND MISC.	76	58	40	198
SUBTOTAL, COLD STRUCTURE	6077	4733	3374	15349
RADIATION SHIELD	76	51	40	114
INSULATION	7	5	4	11
SUSPENSION RODS AND MISC.	30	21	16	45
VACUUM JACKET WALLS	267	236	181	401
SUBTOTAL DEWAR	380	313	241	571
TOTAL WEIGHT	7320	5709	4070	18239

Table 3.4-3. Direct Capital Costs for Magnet * (Millions)

CASE				COLD						
OUTPUT MWe	FIELD TESLA	L m	D_{WB} m	CONDUCTOR $20/kg	STRUCTURE $10/kg	DEWAR $16/kg	FACTORY $20/man-hr	SITE $30/man-hr	MISC COSTS	TOTAL COSTS
1257	6-5	25	5.4	17.3	60.7	6.0	10	10	12	116
887	6-5	20	4.8	13.2	47.3	5.0	8.5	8.5	9	91.5
655	6-5	15	4.1	9.[illegible]	33.7	3.9	7	7	7	67.7
1293	8-7	25	5.4	69.6**	153.5	9.2	22	22	20	296.3

*ESTIMATE PREPARED BY GE-ENERGY SYSTEMS PRODUCTS DEPARTMENT
**$30/KG

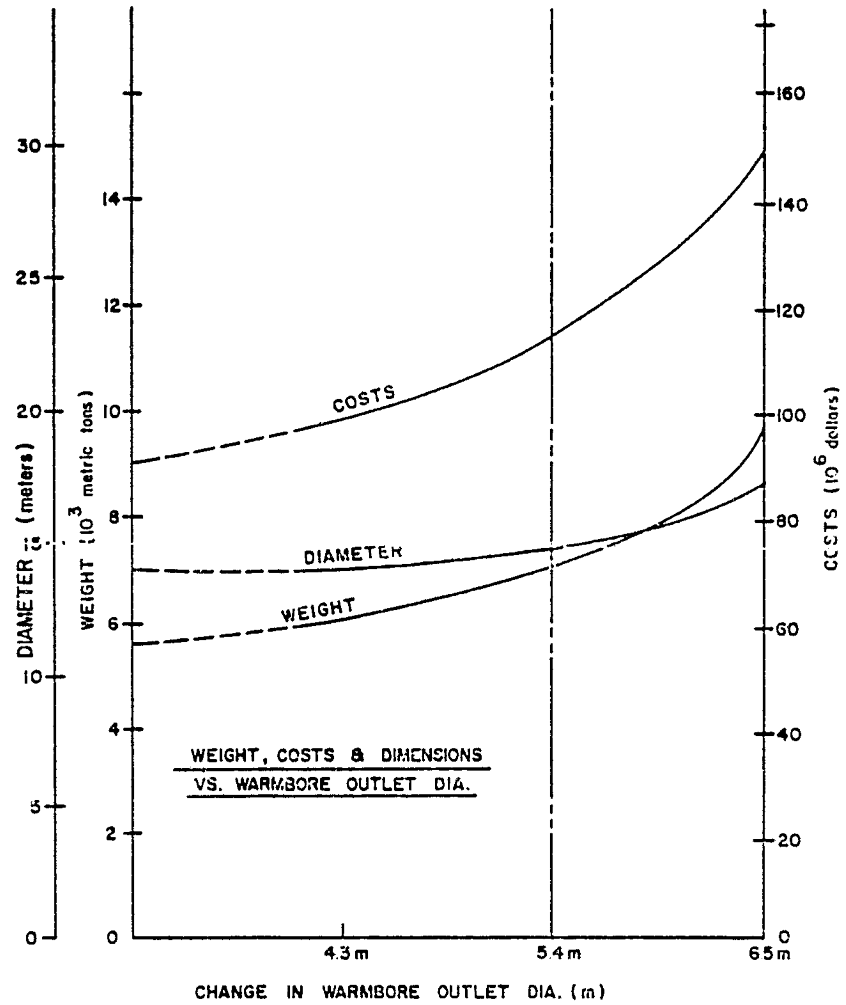


Figure 3.4-3. Magnet Weight, Cost and Outside Diameter as a Function of Warmbore Size (Total Length Constant 29.4m)

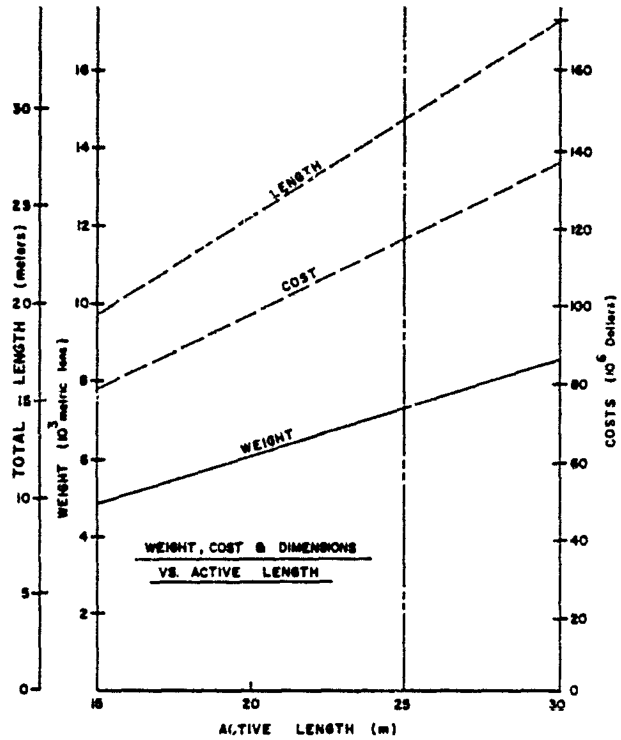


Figure 3.4-4. Magnet Weight, Cost and Total Length as a Function of Active Length (Outside Diameter Constant 15.2m)

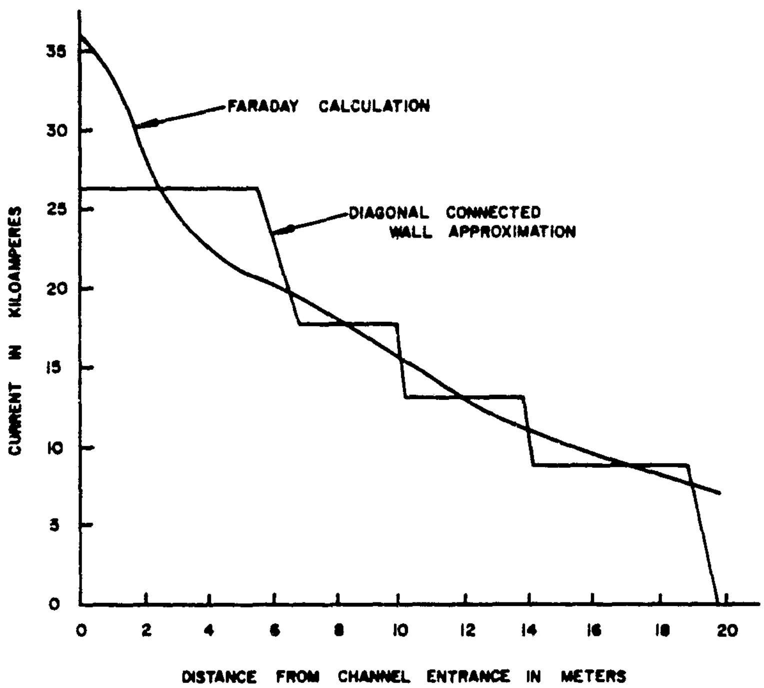


Figure 3.5-1. Current Distribution on Basis of Faraday Generator Analysis Showing Diagonal Connected Wall Approximation Used for Sizing the Power Conditioning System

4400 and 4400 amperes, respectively, to the MHD generator output, while the exit region current of 8800 amperes brings the total current to a nominal 26,400 amperes which must be returned to the plasma in the entrance region. The voltage between adjacent taps is a nominal 9250 volts.

The power conditioning system for this case - through inversion - is illustrated in Figure 3.5-2. Consolidation of the electrode current at each power take-off is accomplished by inverter/converter pairs arranged in a manner similar to that discussed in Appendix X of Reference 1. This arrangement minimizes losses while preserving adequate control of individual electrode current within the take-off region. Inversion is accomplished by a suitable combination of standard current-fed line-commutated twelve-pulse inverter valve modules which are each rated for a nominal 9.25 kV at 4400 amperes. In general, two modules share a single converter transformer in order to improve efficiency and minimize transformer cost. All of the transformer primaries operate from an intermediate 69 kVac bus at which harmonic filters and power factor correction capacitors are switched in as required. A station-level transformer is required to obtain the final transmission level voltage.

Cost estimates for the power conditioning equipment rely heavily on extrapolation from ETF conceptual design studies[1,2,3], HVDC experience and the Westinghouse Electric Corporation study of power conditioning for advanced energy conversion[4]. The breakdown of equipment costs for Case 2.2 is as follows:

Inverter Modules, including system controls	$12.8M
DC Reactors	3.5M
Filters and Power Factor Correction	2.6M
Switchgear, a.c. and d.c.	4.8M
Transformers	11.7M
Current Consolidation Equipment	3.2M
Total	$38.6M

Based on the 585 MW d.c. output from the generator, the power conditioning equipment cost per kilowatt is about $66*. This is considerably higher than for a comparably rated HVDC terminal, primarily because of the use of a relatively large number of inverter modules and transformers operated at relatively low voltage and high current and the need for a significantly more complex instrumentation and control subsystem. This $66/KW cost has been applied to all of the cases to determine power conditioning cost. Though sufficient differences exist among the cases to dictate different power take-off and inverter configurations, overall cost per kilowatt differences are too small to be significant to this study.

*Of this $66/KW MHD, $60.50/KW is inverters and $5.50/KW is for voltage consolidation.

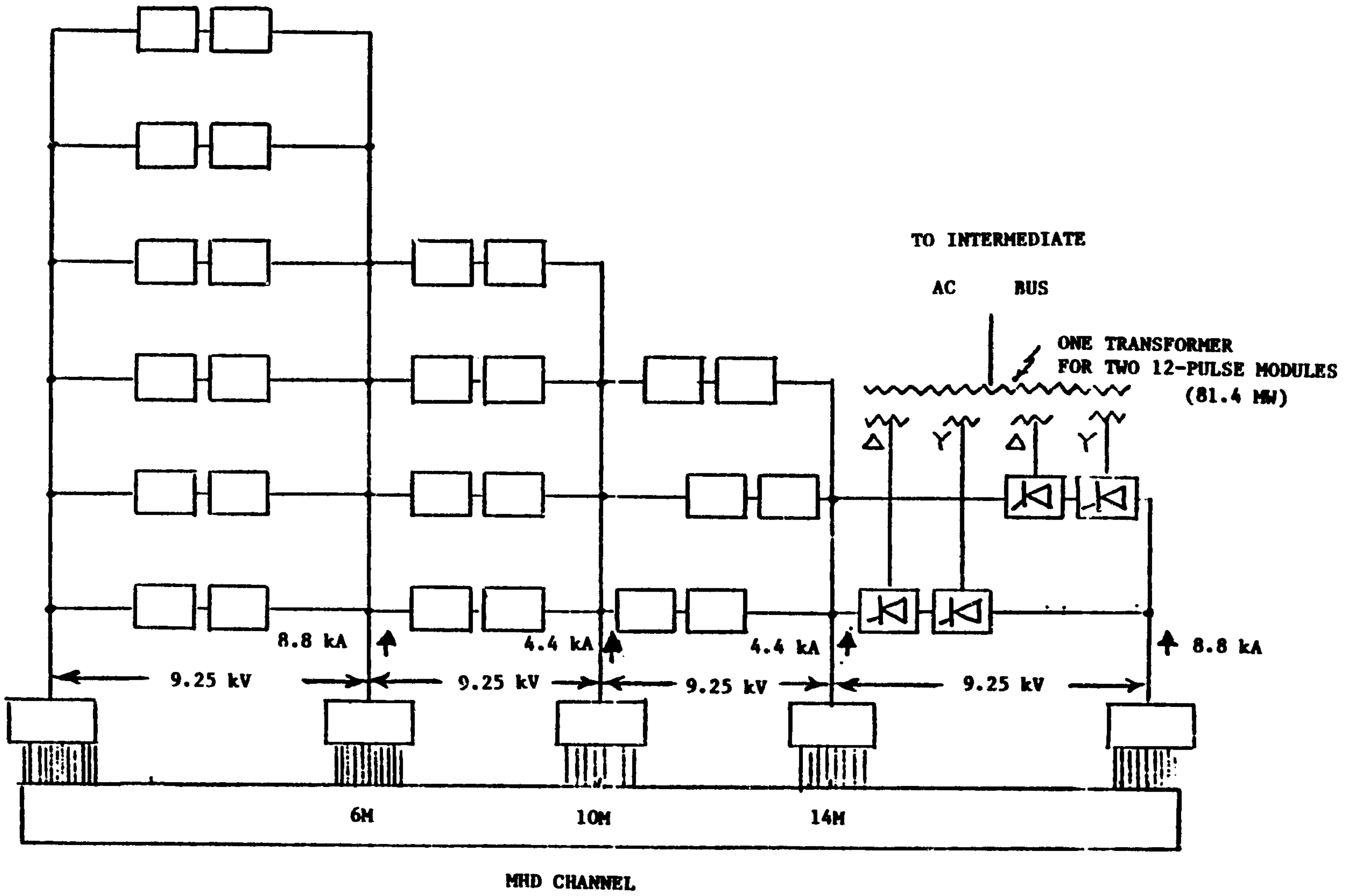


Figure 3.5-2. Typical Power Conditioning System through Inverters Assuming Diagonally Connected Wall MHD Generator with 5 Power Take-Off Locations. Data Based on Case 2.2.

3.6 DIFFUSER

3.6.1 PRESSURE RECOVERY

In the MHD flowtrain, the function of the diffuser is to recover the kinetic energy of the high velocity plasma as it is slowed down to a speed more amenable to the HRSR portion of the system. While subsonic diffuser performance has been widely investigated, both analytically and experimentally, the very high boundary layer blockages expected at the MHD generator outlet present a unique design challenge not usually addressed. The computational results of Doss[1], however, do establish the trends.

From a series of tests run at high blockage[2], the primary conclusions are that the divergence half-angle should not exceed 2 degrees and the pressure recovery coefficient is approximately 48%. The computations of Doss have been normalized with respect to the experimental data and have been used to evaluate diffuser performance as a function of its length-to-initial-width ratio.

In this parametric study, for purposes of generator analysis, diffuser pressure recovery was held constant at 0.60. As shown in Figure 3.6-1, this requires a relatively long diffuser, or active control of the boundary layer which is difficult in a slag/seed laden flow. In his analysis, Doss encountered boundary layer separation prior to a pressure recovery of 0.60 with the presumption that separation takes place at the point where numerical singularities develop. However, the point of three-dimensional boundary layer separation is very difficult to predict and until more data becomes available for high blockage flows at increased lengths, the assumed pressure recovery is a reasonable extrapolation of current data.

A generator performance calculation, done for Case 1.2 (25 meter channel and a magnetic of 6 Tesla tapering to 5 Tesla) with $C_p = 0.5$ resulted in a gross MHD generator output of 567.9 MW. This is a reduction of 12.4 MW from the result with $C_p = 0.6$ and, on the basis of the correlation between enthalpy extraction and plant efficiency, Figure 2.2-6, results in a decrease in efficiency of 0.37% for Case 1.2.

3.6.2 HEAT TRANSFER

The determination of the diffuser heat load consists of evaluating the convective and radiative heat fluxes, summing them, and integrating them over the diffuser wall surface. The convective heat flux is fixed by the Stanton number as derived from a modified form of the Reynolds Analogy. Velocity reduction is treated via "Simple Area Change." Gas properties are evaluated at the Eckert reference temperature from a table of values generated by the equilibrium thermodynamic description of the coal/air/seed combustion product working fluid. The radiative flux is assumed to be due to the presence of CO_2, H_2O, and CO in the gas. The geometry of the gas mass, the gas temperature, the wall temperature and the gas constituent partial pressures characterize the plasma so that Hottel's[3] well known charts of gas emissivities and correction factors can be used.

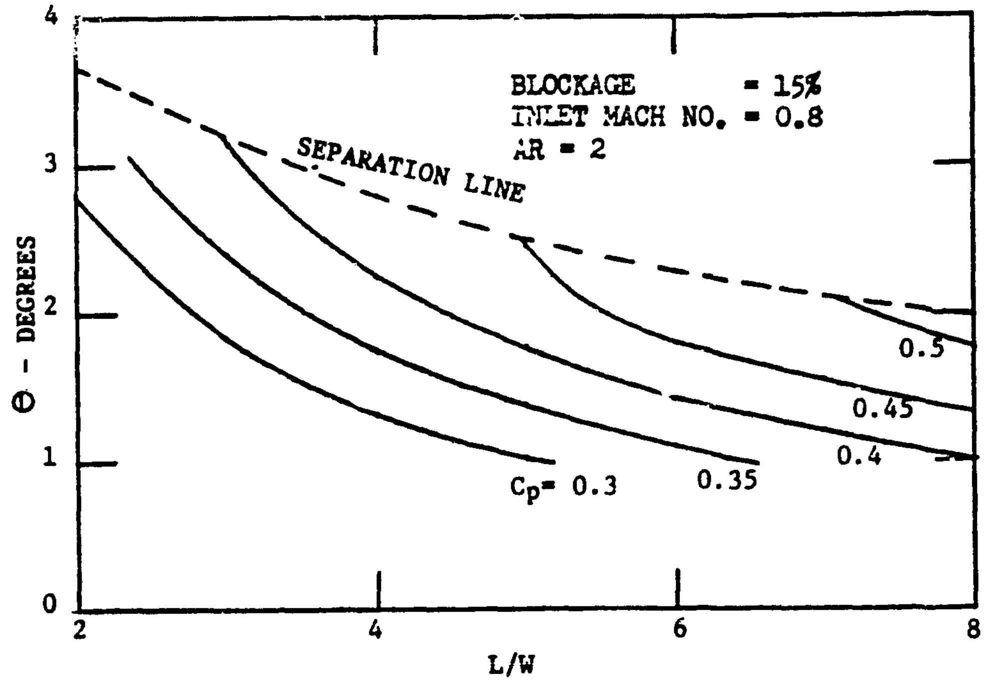


Figure 3.6-1. Diffuser Half-Angle vs Diffuser Length for Various Pressure Recovery Coefficients (Reference 1)

3.7 HEAT RECOVERY/SEED RECOVERY SUBSYSTEM AND PREHEAT COMBUSTOR

The heat recovery/seed recovery (HRSR) subsystem is the group of components in the MHD flow train between the MHD diffuser exit and the stack, e.g., the bottoming plant. The functions of this equipment are:

1. Heat recovery (steam generation and air preheat),

2. Seed recovery, in a manner which permits efficient seed reprocessing, and

3. Environmental control of gaseous and particulate emissions.

The HRSR subsystem configuration adopted in the present study is similar to the concept evolved by Combustion Engineering for the AVCO/C-E/ETF. It is appropriate to a moderate slag carryover system and is a result of applying engineering design practice for chemical heat recovery boilers[1,2] (Kraft boilers) to the HRSR subsystem proposed in ECAS II[3].

The bottoming plant, Figure 3.7-1, consists of four chambers: the radiant furnace, the final oxidation furnace, the convective pass and the backpass.

The gas entering the radiant furnace is cooled slowly, over about a 2-second interval, from ~3500 F to 2900 F, where the gas energy released is transferred to steam in the waterwalls of the furnace. As the flow passes through this furnace, the NO_x concentration is reduced to below EPA limits and vaporized coal slag condenses and begins solidification.

The gas then enters the final oxidation furnace where energy is extracted by superheater tube panels and by steam in the waterwalls of the furnace. Air is injected through ports in the side walls for final combustion of unburned hydrocarbons, carbon and carbon monoxide. The gas temperature at the bottom of the final oxidation furnace is about 1600 - 1800 F. In this temperature range, most of the potassium seed has been converted to solid potassium sulfate, effectively removing the coal-derived sulfur. Part of this seed compound is collected at the bottom of the furnace. Most of the rest passes through the remaining HRSR subsystem to be captured in the stack-gas cleanup equipment (electrostatic precipitator, ESP).

The convective and back passes are filled with metal tube bundles which are used for air preheat, initial steam superheat, steam reheat and feedwater heating. In contract to a conventional coal-fired steam plant, where many of these tube bundles are horizontal, a vertical configuration with very wide tube spacing is used for the HRSR subsystem to prevent plugging of tube passages by the seed-laden gas.

Environmental control is thus accomplished by NO_x reduction in the radiant furnace and secondary combustion in the final oxidation furnace, sulfur capture by the seed in the final oxidation furnace, and particulate removal by the ESP. Heat recovery involves transfer of all of the gas energy, from the radiant furnace inlet to the stack, into water, steam and air. Seed recovery occurs via condensation of potassium sulfate and subsequent collection in the final oxidation furnace, tube banks and ESP.

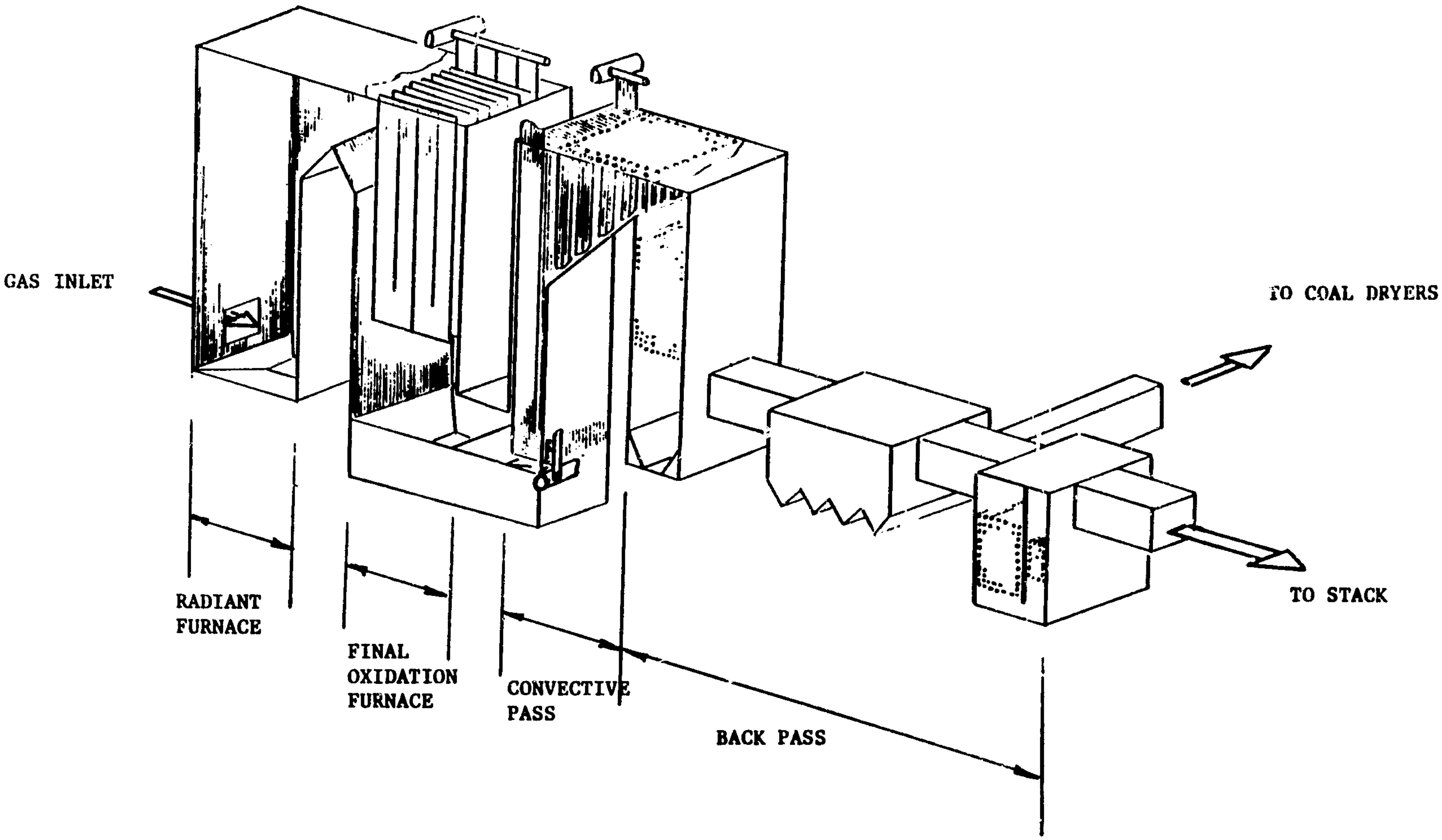


Figure 3.7-1. HRSR Subsystem Configuration

3.7.1 GAS-SIDE ARRANGEMENT

The gas-side circuit, with key temperatures, is shown in Figure 3.7-2.

The radiant furnace is used for NO_x control and slag condensation and rejection. Gas enters the furnace from the diffuser in the temperature range of 3400 - 3600 F, and is cooled slowly to 2900 F. A gas residence time of ~2 seconds is sufficient to permit the NO_x to decompose from a concentration of several thousand ppm at the inlet to below EPA standards (~500 ppm) at the exit. The 2900 F exit temperature is dictated by the fact that the NO_x kinetics are effectively frozen below this temperature, but the seed remains in the gas phase.

A secondary function of the radiant furnace is to reject as much slag as possible. In the temperature range occurring in the furnace, the slag condenses and begins solidification, whereas the seed remians in the vapor phase. Therefore, to minimize mixing of the seed and slag in the downstream components (and thus possibly some loss of the seed), it is advantageous to reject as much slag as possible in the radiant furnace. In the configuration shown in Figure 3.7-1, an active slag removal technique (such as a slag screen) is not used. Rather, slag rejection occurs via the gas fluid mechanics, relying on the inlet jet and turbulence in the bottom of the furnace to deposit wet slag on the furnace walls. It is estimated (Reference B7) that ~40% of the slag carryover will be removed by this mechanism.

The gas temperature drop in the final oxidation furnace is from 2900 F to below 2000 F. As the gas is cooled through this temperature range, the seed condenses, primarily via the exothermic oxidation reactions

$$2KOH(g) + SO_2(g) \rightarrow K_2SO_4(c) + H_2, \tag{1}$$

and

$$H_2 + 1/2O_2 \rightarrow H_2O, \tag{2}$$

where (g) and (c) denote gas and condensed phases, respectively. The dew point, or temperature at which initial condensation occurs, is a function of the gas equivalence ratio (ER) and is in the range of 2100 - 2500 F for $0.90 < ER^{-1} < 1.05$ and typical seed concentrations. Complete condensation of all of the seed occurs over a temperature interval of 200 - 300 F below the dew point. After condensation, the seed is carried along in the flow as suspended droplets. These droplets subsequently solidify in the temperature range of 1800 - 1900 F, which occurs in the bottom of the furnace.

Some seed rejection occurs in the furnace, primarily on the array of superheater panels in the top of the furnace. Here, the typical metal temperature is about 1100 F, which is below the seed condensation point, thus causing the seed to condense on these surfaces and to subsequently drop to the bottom of the furnace. There will also be a small amount of seed condensation on the furnace walls (which also occurs in the radiant furnace) and some of the suspended seed particulates will drop out of the flow at the bottom of the furnace, as the flow turns and exits horizontally from the unit.

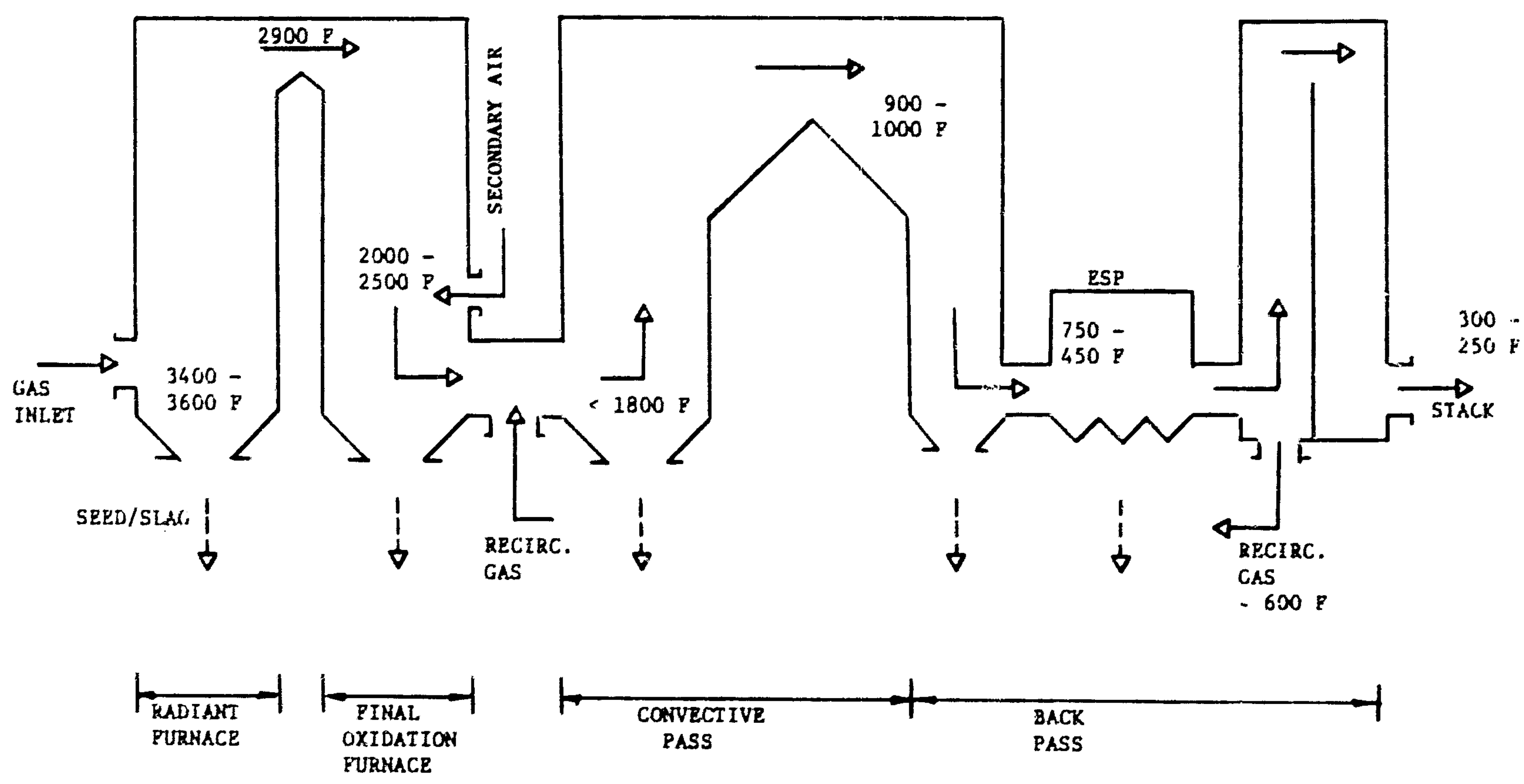


Figure 3.7-2. Gas-Side Circuit HRSR Subsystem

In addition to the seed phenomena occurring in the final oxidation furnace, secondary air is introduced in the lower half of the furnace to complete oxidation ($ER^{-1} = 1.05$) of the fuel-rich combustion gas ($ER^{-1} = 1.05$) of the fuel-rich combustion gas ($ER^{-1} = 0.9$) that enters the furnace. The final oxidation must be conducted in the temperature range of 2000 - 2500 F, the upper limit dictated by secondary NO_x formation and the lower limit by the rate of CO oxidation. Because of the sensitivity of seed dew point to equivalence ratio, the seed condensation is intimately tied to final oxidation.

The gas temperature entering the convective pass should be below the solidification temperature of the suspended seed particulates (< ~1800 F). This procedure insures that the seed is "dry," thus minimizing accumulation of the seed on the tube banks and potential plugging of the tube spaces. To establish this gas temperature, flue gas is recirculated from the gas/steam pinch-point (the high temperature economizer outlet) back to the connecting duct between the final oxidation furnace and the convective pass. In this way, the temperature of the gas leaving the furnace can be adjusted to insure that the seed is dry.

In the convective pass, the gas is cooled by transferring energy to air and steam via various arrays of tube banks. Some seed (and ash) will deposit on the heat transfer surfaces, and is dislodged by sootblowing equipment. The exit temperature from the convective pass, 900 - 1000 F, is the point at which the transition is made from a waterwall enclosure to an adiabatic wall enclosure.

From the back pass inlet, the gas is cooled in additional tube banks to a temperature in the range of 450 - 750 F, at which point it is admitted to the electrostatic precipitator (ESP). The upper temperature limit is characteristic of the maximum operating temperature for state-of-the-art ESP's. Below ~450 F, the seed (K_2SO_4) begins to pick up moisture, forming $KHSO_4$ (Reference 4) a situation undesirable for seed reprocessing. In the ESP, the seed and ash remaining in the flow are removed. The cleaned gas is subsequently cooled through a final set of tube banks to reduce it to stack temperature (250 - 300 F).

3.7.2 STEAM-SIDE ARRANGEMENT

The steam-side circuit for the HRSR subsystem is shown in Figure 3.7-3. A supercritical steam cycle was used for all cases, but comment is made below on the differences in steam-side arrangement for a subcritical steam cycle.

In the steam cycle arrangement used in the present study, water exit the feedwater heater train at 510 F. The low temperature economizer (the downstream economizer in Figure 3.7-3) and MHD channel cooling are interleaved into the feedwater heater train. The water, coming from the last stage of feedwater heating at 510 F, is passed through the high temperature economizer (the upstream economizer in Figure 3.7-3) and then the combustors, nozzle and diffuser. This heat addition raises its temperature to slightly below the critical temperature (705 F), typically 675 - 700 F.

At this point, the water is introduced into the waterwall of the convective section (noted as "from diffuser" in Figure 3.7-3). In this region of modest heat flux, the water is passed through the transition region, exiting the section at 720 F. The enthalpy added to the steam in this pass is analagous to evaporation and initial superheat duty in a subcritical steam cycle.

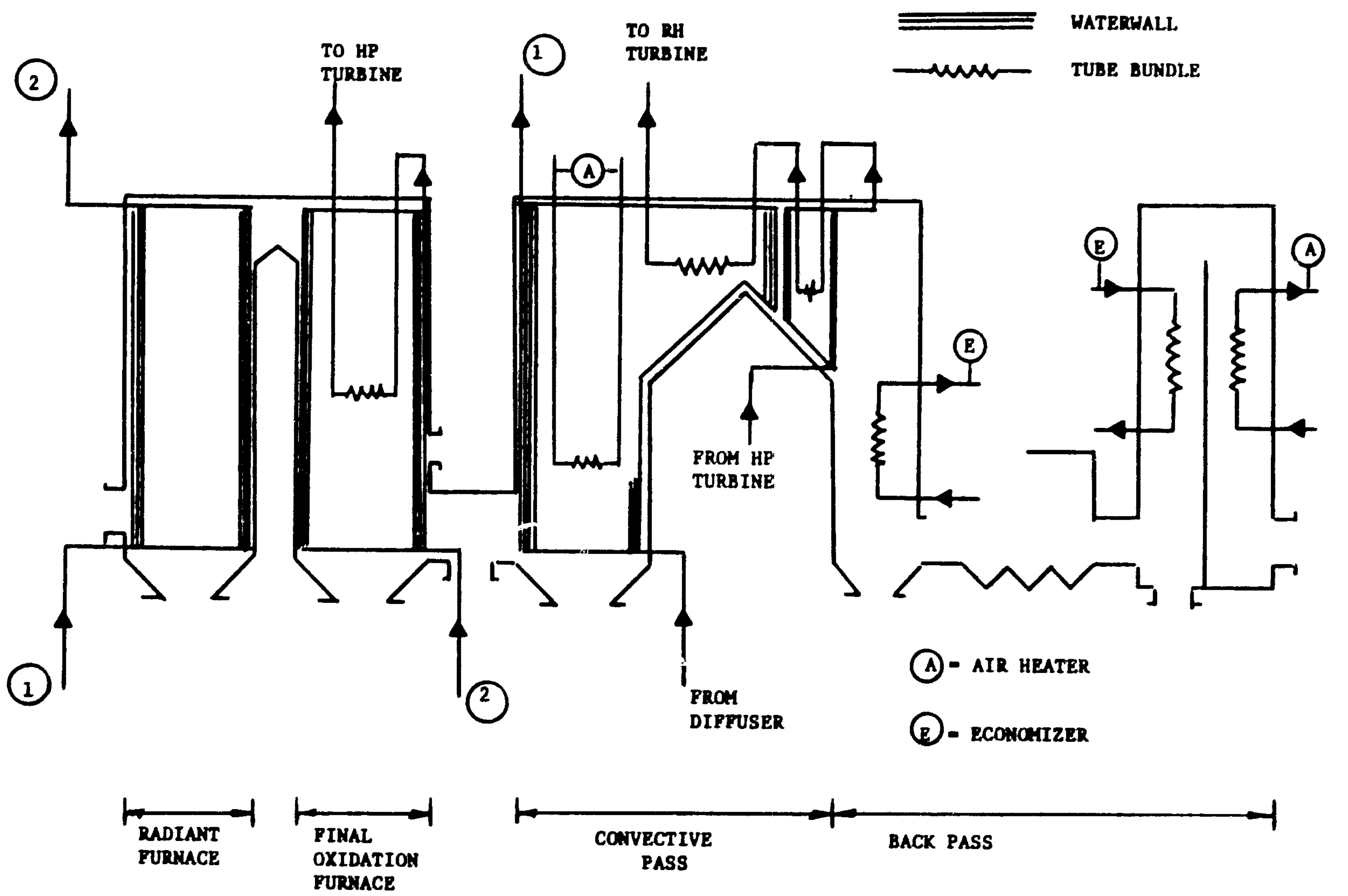


Figure 3.7-3. Steam-Side Circuit (Supercritical) HRSR Subsystem

From the convective pass, the steam is successively passed through the waterwalls of first the radiant furnace and then the final oxidation furnace for intermediate superheat, raising its temperature to 800 - 850 F. Final superheat, to the high pressure turbine inlet temperature of 1000 F, is done in the panels suspended in the upper region of the final oxidation furnace (See Figure 3.7-1).

Steam is returned from the high pressure turbine exhaust at ~600 F to the low temperature reheater, which consists of waterwall and tube bundles in the rear of the convective pass. Here, its temperature is raised to 700 F, and it is then passed through the intermediate and high temperature reheaters, raising its temperature to 1000 F before admission to the reheat turbine. The intermediate and high temperature reheaters are the reheat tube bundles in Figure 3.7-3 that are enclosed by the transition region waterwall.

In a typical subcritical steam cycle (1000 F/1000 F/2500 psi), the modification to the arrangement shown in Figure 3.7-3 would be that the waterwall of the convective pass and the radiant furnace would be used for evaporator duty, the waterwall of the final oxidation furnace for intermediate superheat, and as presently used, the panels in the final oxidation furnace for final superheat. The gas-side temperature distribution would be approximately the same for either case.

3.7.3 RADIANT FURNACE SIZING

For the radiant furnace heat transfer model, the following assumptions were employed:

1. The temperature of the gas flowing through the radiant furnace is reduced from 2300 K (3680 F) to 1900 K (2960 F). This temperature range assumes reasonable diffuser exit conditions, and extends down to the freezing limit of the oxides of nitrogen (NO_x) in the gas stream.

2. The radiant furnace operates at atmospheric pressure. A balanced draft furnace is desirable from an operational viewpoint and is more or less required by the structure of a radiant furnace. With the gas side at atmospheric pressure, the gas density is independent of mass flow rate. Thus, for a specified furnace cross-section, a change in the mass flow rate of the gas will result in a change in the flow velocity but not the density.

3. The convective heat flux in the radiant furnace is negligible. For large radiant furnaces with reasonable wall temperatures, and with gas temperatures as in assumption 1, radiation is by far the dominant heat transfer mechanism. Thus, the local thermal flux density is independent of gas velocity and, because of assumption 2, independent of mass flow rate.

4. The radiant heat flux in the furnace is purely radial. This assumption is analytically convenient, though strictly valid only for furnaces with very large aspect ratios and no axial thermal gradients. For the furnaces considered in this analysis, the aspect ratio was defined as the ratio of the furnace height to the furnace width (square cross-section). An aspect ratio

of 3 to 1 is considered a reasonable minimum value for design purposes and should be sufficient to satisfy this assumption.

5. Under steady-state conditions, an unlined furnace water wall will operate at a slag melting temperature of approximately 1700 K (2600 F) for moderate slag carryover systems. This assumption presumes that a condensate layer will build up on the furnace wall until liquid runoff is established. The condensate layer provides a lower bound on the interior wall surface temperature, independent of the water/steam side conditions. The temperature of 1700 K was chosen as representative of the system design conditions. Precise temperature values are not essential to the radiation cooling calculations.

3.7.3.1 Heat Transfer Analysis

The radiation analysis was performed using the method of Hottel and Sarofim[5], with a correction for particulate emission. Gaseous emission is assumed to be primarily due to the CO_2, H_2O and CO in the gas stream. The equilibrium thermodynamic description (CCE Code), Appendix C, of the coal/air/seed/combustion product mixture was used to generate tables of the thermodynamic properties of the above constituents. These property tables, combined with the gas temperature, the furnace wall temperature and a recommended mean beam length, sufficiently characterize the gas stream that Hottel's well-known charts of gas emissivities and correction factors can be used.

Particulate radiation is treated empirically in the manner of Bueters[6] by using the factor F_E as described in the AVCO ETF Report[7]. This particulate radiation correction factor is applied to both emission and absorption. A value of $F_E = 1.1$ was assumed, but there is considerable uncertainty as to the actual effects of the particulate radiation. Due to this large uncertainty, further complications arising from such considerations as the distinction between slag and soot particles and the depression of the particle temperature below the gas temperature have been ignored in this analysis.

The convective heat transfer was computed using a modified form of the Reynold's analogy which reduces the Stanton number to a function of the local Reynold's number only for fully turbulent flow over a flat plate.

Calculations showed that the assumption of negligible convective heat flux is valid. Of the total heat transferred in the radiant furnace, the contribution due to convection, in all cases considered, was less than 5%. As a result of this condition, the local thermal flux density in the radiant furnace is independent of the mass flow rate of the gas, and becomes solely a function of the gas temperature, the wall temperature and the furnace width.

The results of the heat transfer calculations are shown in Figure 3.7-4. With the wall surface temperature equal to the slag melting temperature of 1700 K, there is a minimum residence time required to cool the gas through its specified temperature range associated with a given furnace width. This figure indicates the furnace width (ℓ) plotted as a function of the gas residence time for the assumed gas temperature change through the radiant furnace.

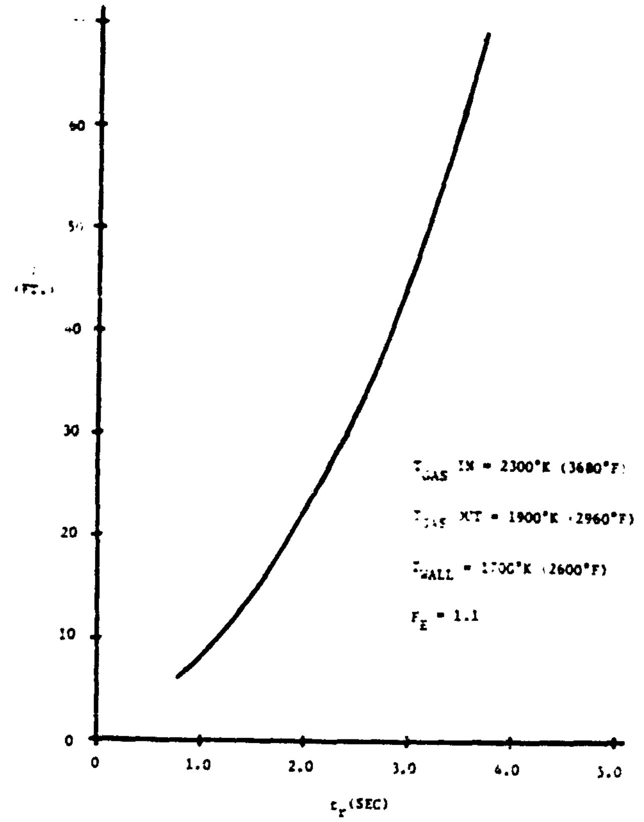


Figure 3.7-4. Furnace Width (Square Cross-Section) as a Function of Gas Residence Time for Cooling of MHD Exhaust Gas with Constant Temperature, Slag Coated Walls.

3.7.3.2 Influence of Mass Flow Rate

To satisfy the need for a reasonable furnace aspect ratio, as noted in assumption 4, it is necessary to consider the mass flow rate of the gas traveling through the furnace. The aspect ratio (n) for this analysis is defined as the ratio of the furnace height (h) to the furnace width (ℓ).

$$n = \frac{h}{\ell} \qquad (2)$$

The velocity of the gas (v) flowing through the furnace is, by definition, equal to the distance traveled divided by the time. Thus,

$$v = \frac{h}{t_r} = \frac{n\ell}{t_r} \qquad (3)$$

The mass flow rate is related to the velocity by the expression

$$\dot{m} = \rho v \ell^2 \qquad (4)$$

which, after substituting for the velocity (v), yields

$$\dot{m} = \frac{\rho n \ell^3}{t_r} \tag{5}$$

An average gas density of $\rho = 0.2$ kg/m^3 for the temperature range considered in this analysis, was selected on the basis of thermodynamic data from the CCE results. The relationship of Equation 5 was used to plot curves of constant $\dot{m}/n$ as a function of furnace width and gas residence time. These curves are shown in Figure 3.7-5. They indicate the furnace width required to obtain a given gas residence time for a fixed aspect ratio and a given mass flow rate.

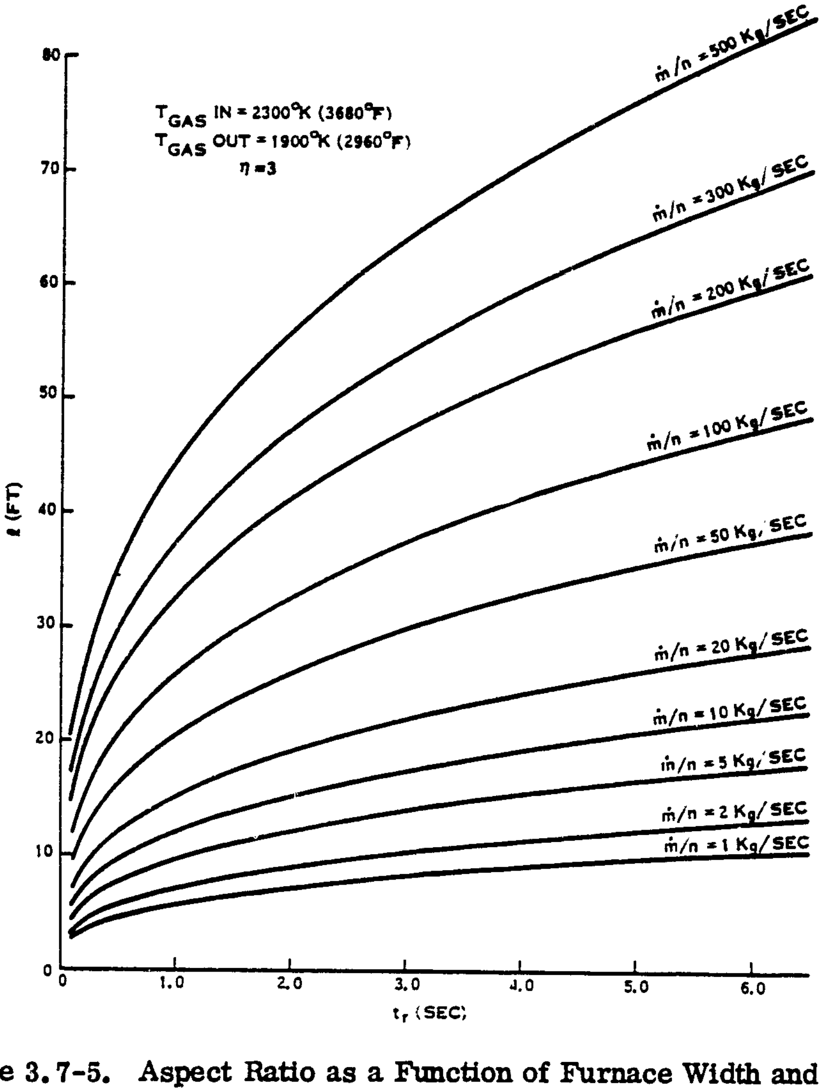


Figure 3.7-5. Aspect Ratio as a Function of Furnace Width and Gas Residence Time

3.7.3.3 Thermal Power Approximation

To complete the analysis, the size of the radiant furnace was related to the total thermal power of the MHD Power Plant.

$$P_{th} = \dot{m}(\Delta H) \qquad (6)$$

where P_{th} is the plant thermal power, $\dot{m}$ is the mass flow of the gas, and ΔH is the enthalpy of the combustion products referenced to standard conditions. For the purposes of this analysis, a value of $\Delta H = 4.6 \times 10^6$ J/kg was assumed. This is an approximation relating MHD generator mass flow to plant thermal power and includes an allowance for losses in the indirectly fired air heater combustion system. Substituting the expression for mass flow (Equation 5) into the above relationship gives

$$P_{th} = \frac{\rho n \ell^3 \Delta H}{t_r} \qquad (7)$$

Equation 7 was used to plot curves of constant thermal power (P_{th}) as a function of furnace width and gas residence time, for a fixed aspect ratio of n = 3. These curves are indicated on Figure 3.7-6.

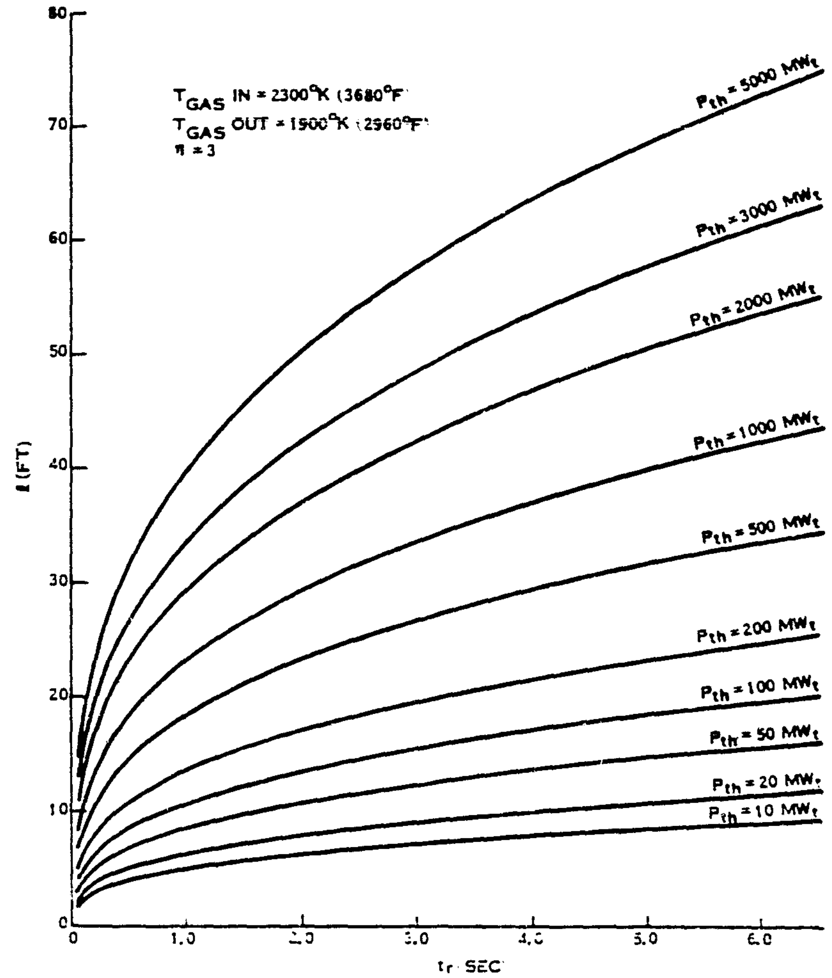


Figure 3.7-6. Thermal Power as a Function of Furnace Width and Gas Residence Time

3.7.3.4 Design Application

Figures 3.7-4 and 3.7-5 describe the relationship between the essential parameters needed in sizing the radiant furnaces for the MHD power plants considered in this study. Their use is demonstrated in the following example of Case 2.0:

Plant Thermal Power, $P_{th} = 2800$ MWt

Mass Flow Rate of Gas, $\dot{m} = 605.6$ kg/sec

Assumed fixed value, $n = 3$

$$\frac{\dot{m}}{n} = 202$$

From Figure 3.7-5, find the curve corresponding to $\dot{m}/n = 202$. Superimposing this curve over the curve shown in Figure 3.7-4, the two lines intersect at the point $\ell = 48.0'$ and $t_r =$ 3.1 seconds. Consequently, for Case 2.0, the correct size of a radiant furnace with a 3 to 1 aspect ratio is 48' wide by 144' high. The time it takes for the gas to travel through this furnace will be 3.1 seconds.

As a quick check on the results of the preceeding example, refer to Figure 3.7-6. Here it can be observed that for a gas residence time of 3.1 sec. and a plant thermal power rating of 2800 MWt, the width of the required furnace having a 3 to 1 aspect ratio is approximately 48 feet.

3.7.4 Design of the Heat Transfer Surfaces

This category includes all of the metallic heat transfer equipment in both the MHD flow train and the preheat combustor flow train. The ceramic, regenerative, high temperature air heaters are discussed separately in Section 3.2.

To establish the heat exchanger tubing requirements for the multitude of cases examined in this study, the procedure used was to complete a conceptual design of the equipment for a single representative case, and then for the remaining cases, to scale surface area and weight based on thermal duty. This procedure assumes that the log-mean temperature difference between the two fluids remains constant in all cases, an assumption which introduces only a modest error (of order 10 - 20%) into the estimates.

The case selected for design was 2.16, since this was the first case for which a final system solution was obtained. Case 2.16 is the same as Case 2.2 (used for BOP and costing by Bechtel) with the exception of MHD channel performance and type of MHD combustor. The heat transfer equipment is identical, within the level of detail of the present study.

The individual heat transfer surfaces considered in the HRSR subsystem are indicated in Figure 3.7-7. In addition, the metallic air heaters in the pre-heat combustor flow train were included in the present analysis (see Sections 2.3.2.1 and 3.1). From the results of the

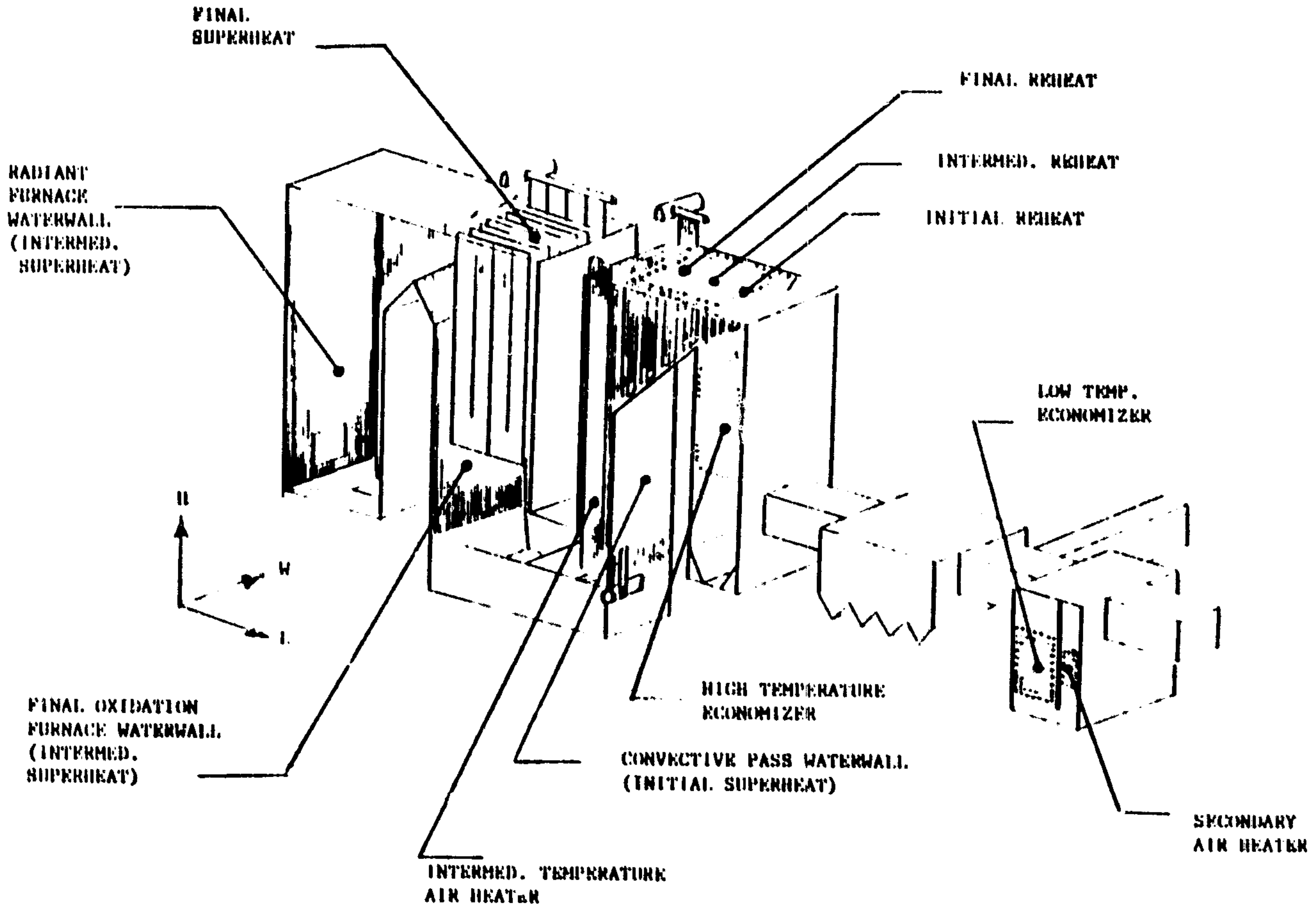


Figure 3.7-7. Heat Transfer Surfaces HRSR Subsystem

system code analysis, flow rates, temperatures, and thermal duties were established (see Table 3.7-1).

Table 3.7-1. Thermal Data for Heat Transfer Equipment, Case 2.16

UNIT	TEMPERATURE (F)				MASS FLOW (KG/S)		DUTY (MW)
	GAS		STEAM/AIR		GAS	STEAM/AIR	
	IN	OUT	IN	OUT			
Radiant Furnace							
• Intermediate suprht	3480	2900	720	740	606	575	288
Final Oxidation Furnace							
• Intermediate suprht	2900	2038	740	770	606/694	575	155
• Final suprht	2900	2340	770	1000	606	575	351
Convective Pass							
• Intermediate Temperature Air Heater	2038	1668	860	1300	694	529	145
• Reheat							
• Final	1668	1489	900	1000	694	510	67
• Intermediate	1489	1127	700	900	694	510	135
• Initial	1127	953	602	700	694	510	79
• Initial suprht	2038	1127	697	720	694	575	94
Back Pass							
• High Economizer	953	612	510	604	694	575	150
• Low Economizer	612	300	190	300	455	485	88
• Secondary Air Heater	300	273	59	200	455	88	7
Preheat Combustor Flow Train							
• Preheat Combustor Air Heater	1503	1187	705	1100	311	281	68
• Low Temperature Air Heater	1187	927	692	860	311	529	54

With the exception of the two furnaces, discussed below, the general procedure entailed the following steps:

1. Select tube configuration and tube spacing.

2. Select gas velocity and calculate gas flow cross-sectional area (width, W, was held constant at that of the two furnaces).

3. Calculate heat transfer coefficients and maximum tube wall temperatures.

4. Select tube material and calculate tube wall thickness.

5. Calculate required transfer area and tube mass.

6. Calculate pressure drops.

7. Iterate as necessary to achieve acceptable pressure drops and heat fluxes.

Some modification to the above procedure, which is appropriate for tube bundles, was required when sizing the waterwall.

3.7.4.1 Heat Recovery/Seed Recovery Design

The radiant furnace was sized as described in Section 3.7.3 and resulted in approximately a 50' x 50' cross-section with a height of 150'. Adjustments from case to case were made as indicated in Section 4. These same dimensions were used for the final oxidation furnace. The final superheat assembly consists of 9 panels of tangentially-welded tubes, each panel 50' x 100', with panels located on 5' centers. Each individual tube is two-passed, with inlet and outlet headers above the furnace.

The air heater assembly in the front of the convective pass also consists of tangentially-welded tube panels, 10' wide by 50' long, located on 21" centers. Here, the individual tubes are single pass, with inlet headers at the top of the unit and outlet headers on the bottom.

The steam reheat assembly consists of waterwall in the back of the convective pass, which accepts the inlet steam from the high pressure turbine exhaust, and then a series of multiple-pass tube bundles to attain the final reheat steam temperature. Waterwall, for initial steam superheat (e.g., the transition region), surrounds the air heater panels and most of the steam reheat bundles.

The three units in the back pass, the high and low economizers and the secondary air heater, are multiple-pass tube bundles with horizontal tube runs. In the present first-order analysis, the tube bundles were treated as bare tubes. However, because of the small temperature driving head in the back pass units, the rather considerable surface area of the economizers could be reduced by the use of finned tubing.

A tabulation of the sizing for the heat transfer surfaces is given in Table 3.7-2.

Table 3.7-2. Specification of Heat Transfer Surfaces, Case 2.16

UNIT	MATERIAL	TUBING				LMTD	U	$\dot{q}''$	A	T_W	M	P				DIMENSIONS
												GAS		STEAM		
		OD	t_w	t_1	t_{11}							IN	Δ	IN	Δ	L x W x H
Radiant Furnace																
Intermediate suprht	SA213/T22	1.5	0.25	1.75	NA	2450	12.3*	30,200*	33*	800	948	14.7	0.10	3850	100	50 x 50 x 150
Final Oxidation Furnace																
Intermediate suprht	SA213/T22	1.5	0.25	1.75	NA	1680	9.7*	16,300*	33*	850	943	14.6	0.15	3750	100	50 x 50 x 150
Final suprht	SA213/TP304h	2.0	0.32	60	2	1730	4.9	8,500	141	1200	2350	14.6	0.10	3650	150	50 x 50 x 100
Convective Pass																
Intermediate Temperature Air Heater	SA213/TP304h	2.0	0.087	21	2	770	5.2	4,000	125	1350	565	14.45	0.08	170	6.5	10 x 50 x 50
Reheat																
Final	SA213/T5	2.0	0.20	6	4	630	7.9	4,980	46	1100	425	14.37	0.03	711	20	5 x 50 x 50
Intermediate	SA213/T2	2.0	0.13	6	4	505	7.9	3,990	115	1000	530	14.34	0.07	748	37	20 x 50 x 40
Inital	SA210/A1	2.0	0.074	6	4	390	7.7	3,[illegible]00	90	800	345	14.27	0.03	768	20	17 x 50 x 25
Initial suprht	SA210/A1	1.5	0.30	1.75	NA	790	9.0*	7,100*	28*	800	825	14.45	0.13	1950	100	Irreg.
Back Pass																
High Economizer	SA210/A1	2	0.31	6	4	200	15	3,000	171	650	2770	14.24	0.17	4220	20	20 x 50 x55
Low Economizer	SA210/A1	2	0.065	4	4	200	13	2,600	116	350	390	14.07	0.05	157	20	23 x 50 x 30
Secondary Air Heater	SA210/A1	2	0.065	4	4	150	10	1,500	16	250	5[illegible]	14.02	0.04	1[illegible].7	0.3	20 x 50 x 5
Preheat Combustor Flow Train																
Preheat Combustor Air Heater	SA213/T22	2	0.10	7.4	2	440	14.9	6,540	36	1175	185	153.9	0.3	172.7	2.3	14 x 14 x 35
Low Temperature Air Heater	SA210/A1	2	0.065	4.7	2	280	15.1	4,240	4[illegible]	910	147	153.6	0.1	171.3	2.4	14 x 14 x 28
Vessel	SA210/A1	NA	3.0**	NA	NA	NA	NA	NA	NA	300**	1470**	NA	NA	NA	NA	100 (L) x 30 (D)

<u>Legend</u>

OD = Tube outside diameter (IN)
t_w = Tube wall thickness (IN)
t_1 = Transverse tube spacing (IN)
t_{11} = Longitudinal tube spacing (IN)
LMTD = Log-mean temperature difference (F)
U = Overall heat transfer coefficient (Btu/hr. ft^2 F)
$\dot{q}''$ = Heat flux (Btu/hr. ft^2)
A = Heat transfer surface area (10^3 ft^2)
T_W = Maximum tube wall temperature (F)
M = Tubing mass (10^3 lb), includes 25% allowance for headers
P = Pressure (psia)
LxWxH = Overall dimensions of unit (FT) - See Figure 5

* = Projected
** = Vessel wall thickness, wall temperature and mass, respectively

3.7.4.2 Preheat Combustor Flow Train Design

A nodal representation of the high temperature air heater (HTAH) system is shown in Figure 3.7-8 (for Case 2.16). The features to note here are, first, that the preheat combustion system is pressurized, and second, that in the combustion gas flow train there are two air heaters below the HTAH. These two air heaters are used to reduce the gas temperature from ~1500 F, at the HTAH outlet, to ~900 F at the gas turbine inlet. This particular turbine inlet gas temperature is required to match the turbine output to the compressor input power requirement.

Because of the reduced air temperatures in the PCAH and LTAH (see Table 3.7-1) as compared to the HTAH, these two units are designed with metallic tubes. The two units are somewhat different than air heaters in the MHD flow train, however, because both of the working fluids are pressurized. Thus, the heat transfer surfaces must be enclosed in a pressure vessel, as is the case with the HTAH.

The conceptual design for the two air heaters is shown in Figure 3.7-9. Both of the air heaters are contained within a single pressure vessel, with the LTAH surfaces above the PCAH surfaces. The heat transfer surfaces are tangentially-welded tube panels, the air making two passes through the gas. The gas enters through the side of the vessel, flows downward and then upward making two passes over the PCAH panels, and then a single pass over the LTAH panels, exiting near the top of the vessel. The specifications for this unit are contained in Table 3.7-2. A number of other cases involved different heat transfer equipment, for example, a water economizer (Case 2) and an atmospheric air heater (Base Case 1 and Case 2.12). The heat exchanger specifications for these other cases were obtained by scaling via thermal duty and log-mean temperature difference.

3.8 SULFUR CLEANUP OTHER THAN SEED CAPTURE

A problem introduced by indirectly-fired high temperature air heaters (HTAH) is cleanup of the (coal-derived) sulfur from the preheat combustion gas. In the present study, four different approaches were examined and two selected for inclusion in case studies. However, none were found to yield a particularly attractive system. The approaches considered were:

1. Coal cleanup,

2. Hot gas cleanup,

3. Flue gas desulfurization

4. Chemically active fluidized bed gasifier.

State-of-the-art physical coal cleaning, based on performance described in Reference 1 and 2, is capable of removing about 1/2 of the sulfur and ash content of the original coal, with an energy penalty of a loss of 5 - 10% of the fuel content. This process was rejected because of inadequate sulfur removal and significant energy penalty. Chemical coal cleaning, Reference 3, has the potential for more thorough and efficient desulfurization. However, this technique is still in the laboratory stage and not well enough defined, from the standpoint of performance

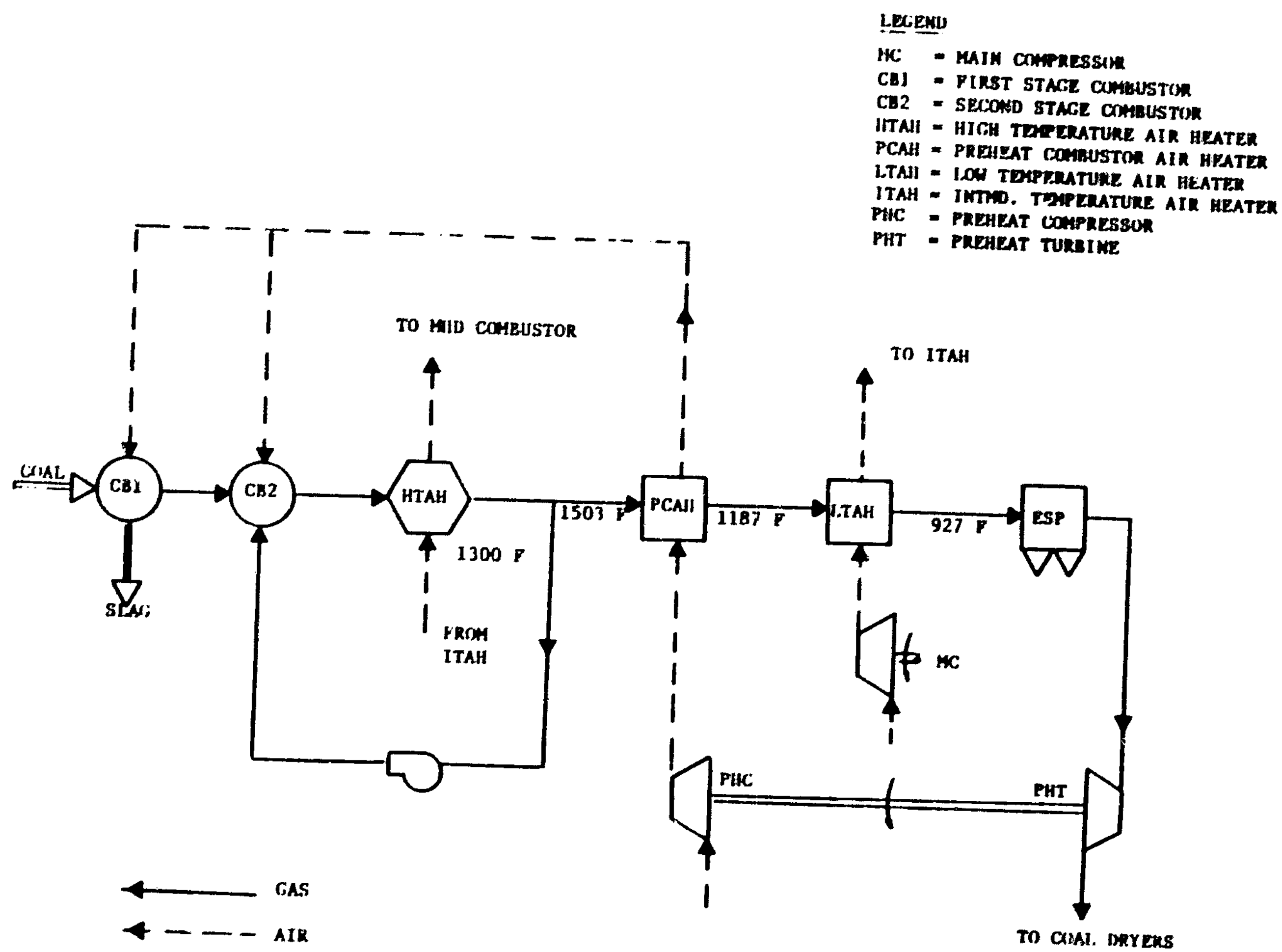


Figure 3.7-8. High Temperature Air Heater System, Case 2.16

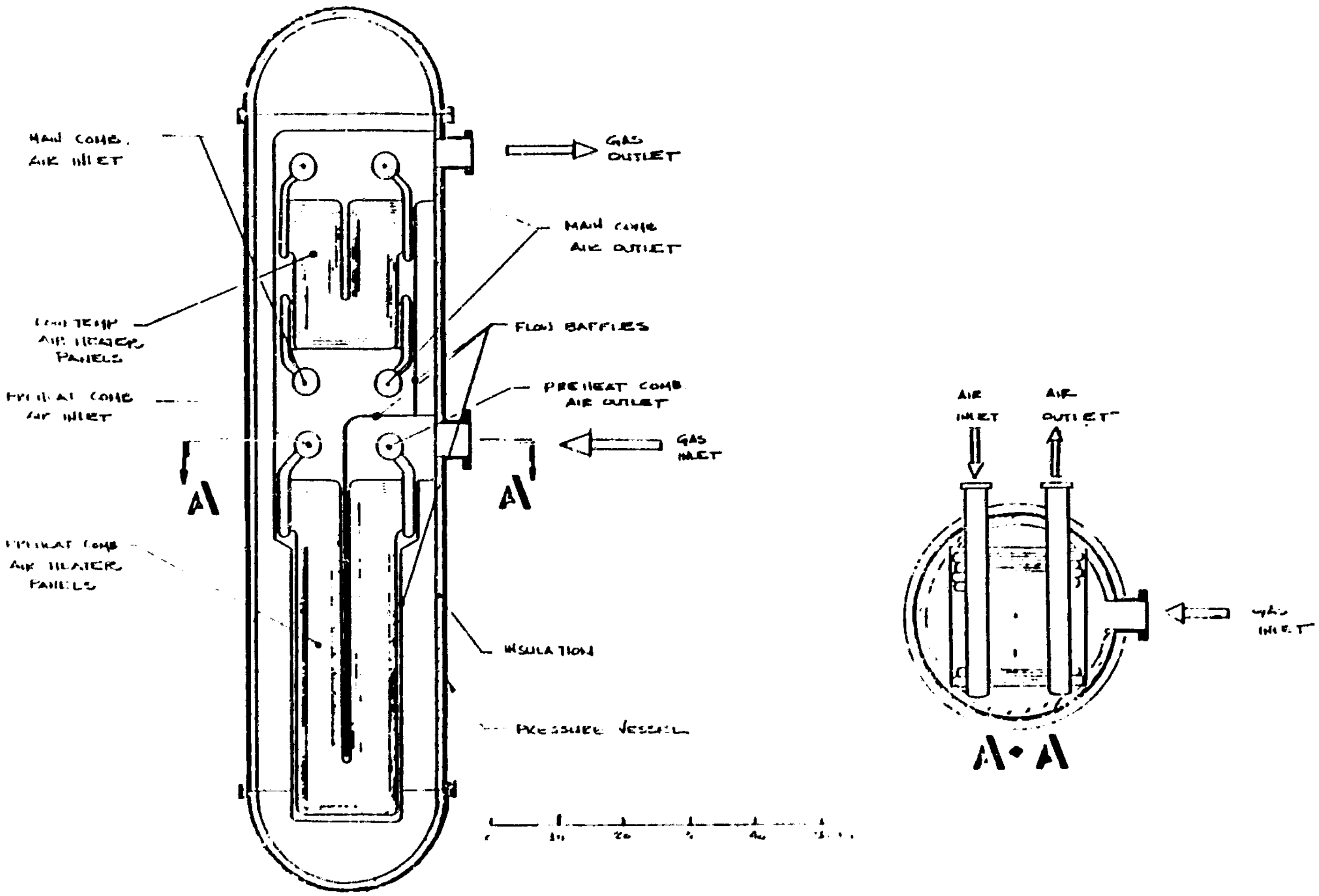


Figure 3.7-9. High Temperature Air Heater System Recuperative Air Heaters

and process details, for consideration in the present study. Although an appropriate coal cleaning technology is not available at the present time, it is likely that this approach to desulfurization, using some combination of physical, chemical (and perhaps other) procedures, will eventually yield the most system-attractive sulfur clean-up technology.

The second desulfurization approach examined was hot gas cleanup, specifically, the iron oxide process being developed by the Bureau of Mines[4] and, independently, by Babcock and Wilcox[5]. The iron oxide process involves use of a regenerable iron oxide bed which removes H_2S from the (hot) fuel-gas stream produced by a gasifier, via the reaction

$$H_2S + FeO \rightarrow FeS + H_2O,$$

with the sulfur being removed from the gas and captured in the bed material. The bed is intermittently regenerated with air, undergoing the reaction

$$\frac{3}{2} O_2 + FeS \rightarrow FeO + SO_2,$$

with the SO_2 being carried off in the regeneration air flow (at a typical concentration of ~10 mole %).

The FeO unit would be installed in the HTAH combustion system as shown in Figure 3.8-1. The highly concentrated SO_2 regeneration stream could be fed into the MHD flow stream for sulfur capture by the seed, or could be sent directly to a Claus plant for reduction to elemental sulfur. Unfortunately, we were not successful in integrating the FeO unit into the system because of the temperature constraints indicated in Figure 3.8-1. The FeO bed must be operated with a gas inlet temperature in the range of 1000 - 1500 F. However, we were not able to find a gasifier (e.g., first stage combustor) configuration with a sufficiently low gas outlet temperature capable of achieving the required 3000 - 3300 F temperature at the second stage combustor exit and this approach was discarded.

The third approach examined, and adopted for all of the systems in base cases 1 and 2 (except case 2.17, the chemically active fluidized bed), is the spray dryer (flue gas desulfurization) system[6,7,8] shown in Figure 3.8-2. An aqueous solution of Na_2CO_3 is sprayed into the flue gas with an atomizer, at a gas temperature of 300 - 1500 F. The atomized droplets of solution mix intimately with the flue gas, allowing the Na_2CO_3 to react with the SO_2 to form Na_2SO_4. Because the liquid-to-gas ratio (L/G) is held low and the water is in the form of fine droplets, the water rapidly evaporates, leaving the reaction products (Na_2SO_4, Na_2SO_3) and unreacted absorbent (Na_2CO_3) as suspended particulates in the gas flow. These particulates, which contain the sulfur, are then removed downstream of the spray dryer with standard gas cleanup equipment (ESP, cyclone or baghouse).

The water requirements for the spray dryer are approximately 3 lb of water/100 lb of flue gas (Reference 7), with the Na_2CO_3 concentration adjustable to suit the SO_2 concentration of the flue gas. The drop in gas temperature through the spray dryer is 125 - 150 F, so that even with a 300 F gas inlet temperature, the gas remains above the water dew point at the exit of the unit. The only energy requirements are electrical power for the atomizer motors

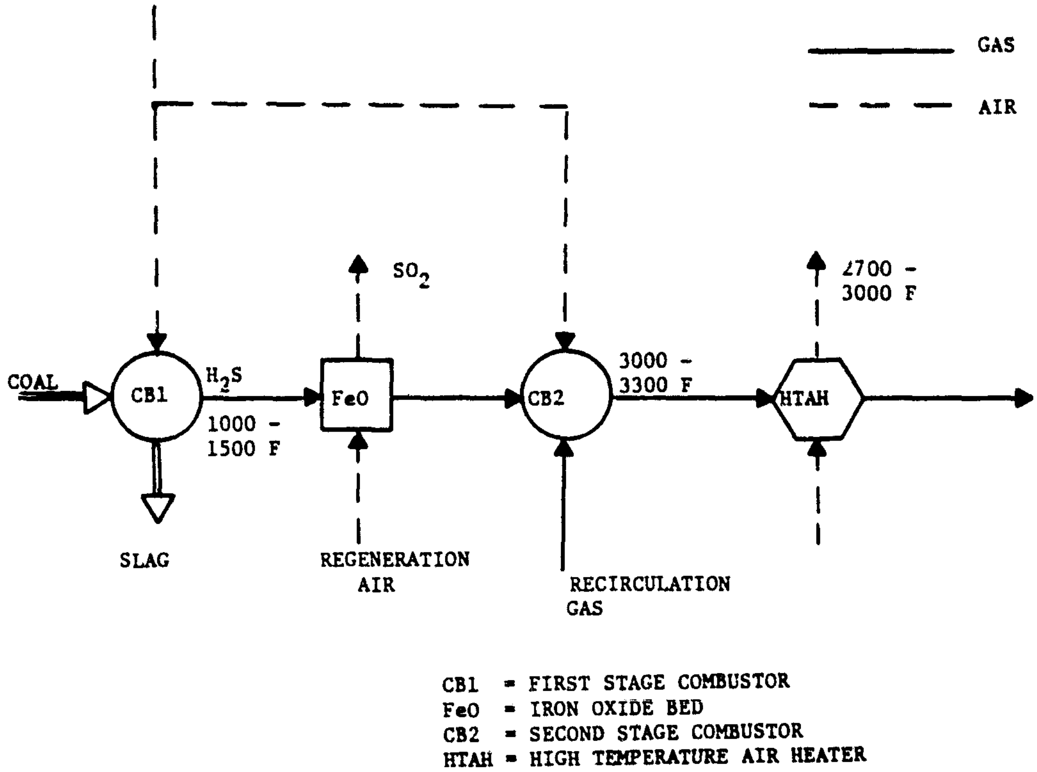


Figure 3.8-1. Iron Oxide Process for Sulfur Cleanup

(neglected here) and a gas pressure drop through the unit of 10 - 15 inches of water.

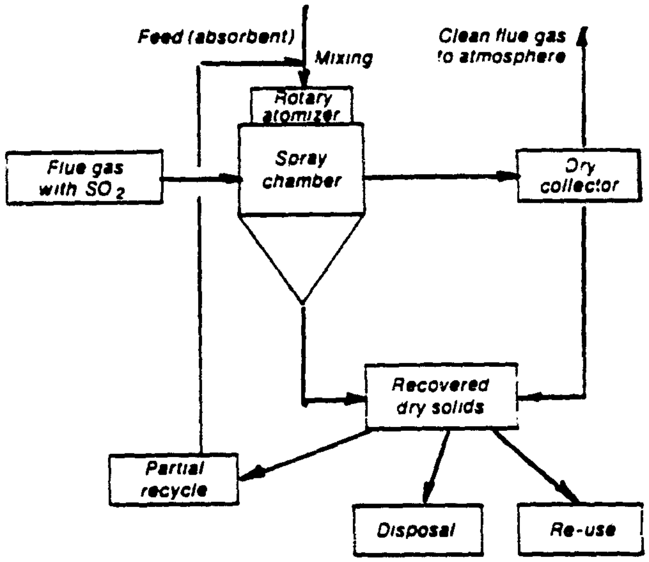
A modification of the spray dryer configuration described above, which has been tested in a pilot facility by Joy/Niro Company (Reference 6), is the use of CaO as the sorbent, rather than Na_2CO_3. The CaO, dissolved in water, forms $Ca(OH)_2$. The hydroxide, when sprayed into the flue gas, undergoes the reaction

$$Ca(OH)_2 + SO_2 + 1/2O_2 \rightarrow CaSO_4 + H_2O$$

with the sulfur thus being captured as $CaSO_4$ particulates rather than Na_2SO_4 particulates. This variation on the spray dryer was used since the seed reprocessing scheme adopted, the formate process, also produces sulfur in the form of $CaSO_4$ as an end product. Thus, all sulfur fired to the combustors is processed into $CaSO_4$.

The chemically active fluidized bed, because it entails considerations other than sulfur cleanup, is discussed elsewhere in this report (Section 3.1.2.7).

Sulfur removal is not without cost to the overall system, both in lowering the plant efficiency and boosting the cost of electricity. For Base Case 1, flue gas desulfurization was used to remove sulfur from the combined preheat and main MHD flow combustion products streams. The wisdom of this choice is seen in Table 3.8-1. Although partial seed reprocessing is a viable



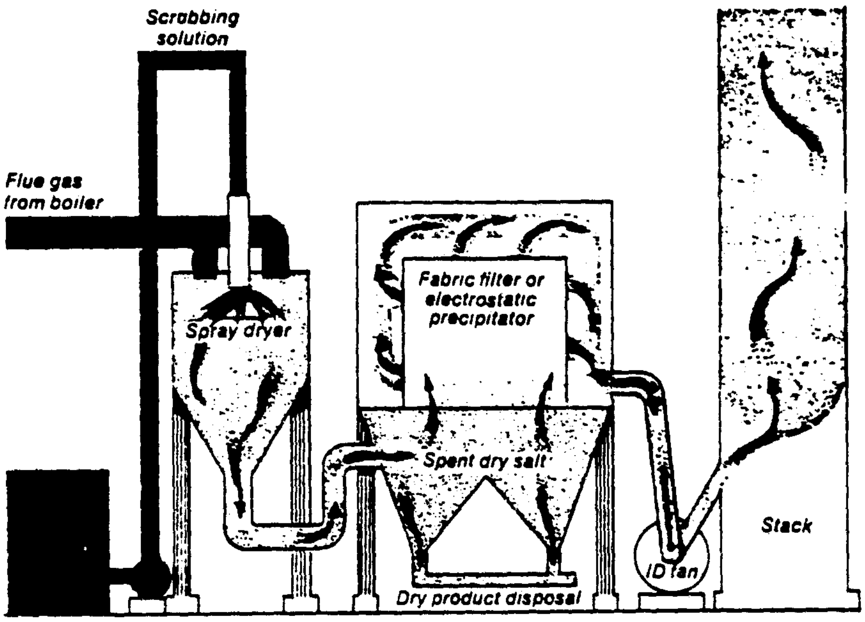


Figure 3.8-2. Spray-Dryer System for Flue Gas Desulfurization (From Reference 6)

alternative, the seed reprocessing plant would have added substantially to the capital and operating costs of the overall plant. Moreover, the reprocessing plant's energy requirements would have cost an additional 0.16 point in overall plant efficiency*. Clearly, the most attractive sulfur cleanup technology will be the one that achieves proper sulfur-removal levels at the lowest cost and at the lowest plant power drain.

Table 3.8-1. The Effect of Sulfur Removal on Case 1.0

SULFUR REMOVAL		DIRECT CAPITAL COST - $$10^6$				OPERATING COST - $$10^6$/YR				EFFICIENCY %
PREHEAT	MHD FLOW	FGD	CGC*	SEED PLT	TOTAL	FGD	CGC*	SEED PLT	TOTAL	
FGD	SEED REPRO.	5.9	-	12.4	18.3	0.6	-	3.9	4.5	41.25
FGD	FGD	14.3	-	-	14.3	1.4	-	-	1.4	41.41
CGC	SEED REPRO.	-	38.7	12.4	51.1	-	5.3	3.9	9.2	39.98

*COLD GAS CLEANUP

*Also appearing in Table 3.8-1 is an estimate of the cost and efficiency penalties associated with a standard cold gas cleanup plant.

3.9 SEED REPROCESSING

3.9.1 STRUCTURE OF SEED REPROCESSING STUDIES

Seed reprocessing has been studied in some detail by the MHD community[1,2,3,4]. However, prior to PSPEC, a major chemical company with actual hands-on experience in the manufacture of potassium salts had never examined the problem. In order to take advantage of chemical manufacturer experience, GE contacted several major chemical companies to explore their interests in a cooperative study. Hooker Chemical Company responded favorably and agreed to consult on seed reprocessing.

The first step in the cooperative effort was to educate Hooker Chemical Company in MHD steam/power generating systems. This was done by the preparation of a position paper (Appendix D) and oral briefing on the seed effluent from the HRSR system. In addition, all documentation available to GE was forwarded to Hooker for their use and information.

Later, a joint meeting between NASA, Hooker Chemical Company and GE was held to discuss the seed reprocessing problem. This meeting led to the conclusion that both electrolytic processing and conversion of potassium sulfate to carbonate using the formate process were promising candidates. Hooker Chemical Company agreed to examine these systems and define preliminary plant layouts, mass and energy balances, plus capital and operating costs.

The study supplied considerable details on both processes under consideration and brought manufacturing expertise to bear on the seed reprocessing problem. Results were incorporated in a letter report included herein as Appendix E.

3.9.2 GE ANALYSIS OF THE HOOKER CHEMICAL COMPANY RESULTS

Examination of the results presented in Appendix E shows that the formate process is superior to electrolytic conversion, both as to capital cost and material expenditures. For this reason, electrolytic processing was discarded and received no further consideration in the present study. The formate process on the other hand, appears realistic and cost effective. The major uncertainty associated with the formate process is the formate reactors themselves. The reactor size which was used to provide capital cost estimates is based on extrapolation from sodium data. A factor of 3 was used to provide conservative estimates for reactor size. Consequently, the capital cost estimates are probably on the high side.

Several aspects of the seed reprocessing study by Hooker Chemical Company did not match the general ground rules for the PSPEC Study. Therefore, the basic data were modified by GE such that the methodology used for costing seed reprocessing agreed with that used in the remainder of the PSPEC study. Specifically:

1. An oxygen plant was costed using the data supplied to GE by NASA.

2. The cost of the coke gasifier was separated from that of the oxygen plant.

3. The mass balances and temperatures supplied by Hooker Chemical Company were used to construct an energy balance for the seed reprocessing system.

4. Pressurized carbon monoxide exiting the formate reactors was burned, diluted with nitrogen and fed to gas turbines which supply shaft power required for the dissolver and formate reactors.

3.9.3 RESULTS OF THE SEED REPROCESSING STUDY

The optimized plant layout for the formate process is shown in Figure 3.9-1.and the mass and energy balances for the system are given in Table 3.9-1. Several comments regarding this system are appropriate:

1. The oxygen plant is assumed to be identical in construction to those used throughout the PSPEC study, but substantially smaller (95 tons/day).

2. A scrubber is required to remove CO_2 and H_2S from the raw gas produced by the gasifier. CO_2 must be removed to prevent precipitation of $CaCO_3$ in the formate reactors. H_2S must be removed from environmental considerations.

3. The size of the formate reactors required in the process is the major uncertainty associated with the seed reprocessing system.

4. The formate solution fed to the flaker is very concentrated. This minimizes the energy required to recover the formate in a solid form and is a significant improvement suggested by Hooker Chemical Company.

5. The excess pressurized CO coming from the formate reactor is fed to a combustor and burned as described in the previous section.

Energy requirements for the seed reprocessing system are supplied almost entirely by the coke fed to the gasifier. In an actual plant the calculated additional shaft power requirement (0.6 MW) would be supplied by feeding a slight excess of coke to the gasifier such that the overall seed reprocessing system becomes thermoneutral.

The seed reprocessing system as designed produces potassium formate. The formate can be converted to potassium carbonate, but this conversion would impose an energy penalty. An alternative is to seed the MHD combustion gases with formate directly. The formate burns to the carbonate in the main MHD combustor and the energy of combustion (0.73 MJ/Kg KCOOH) is recovered in the high temperature section of the MHD/steam power generating system. For plant performance calculations, a debit was taken for the energy in the coke and a credit was assumed for the energy of the dormate conversion to carbonate. The net thermal difference, plus a small penalty required for excess shaft power, gives the energy requirements for the seed reprocessing system.

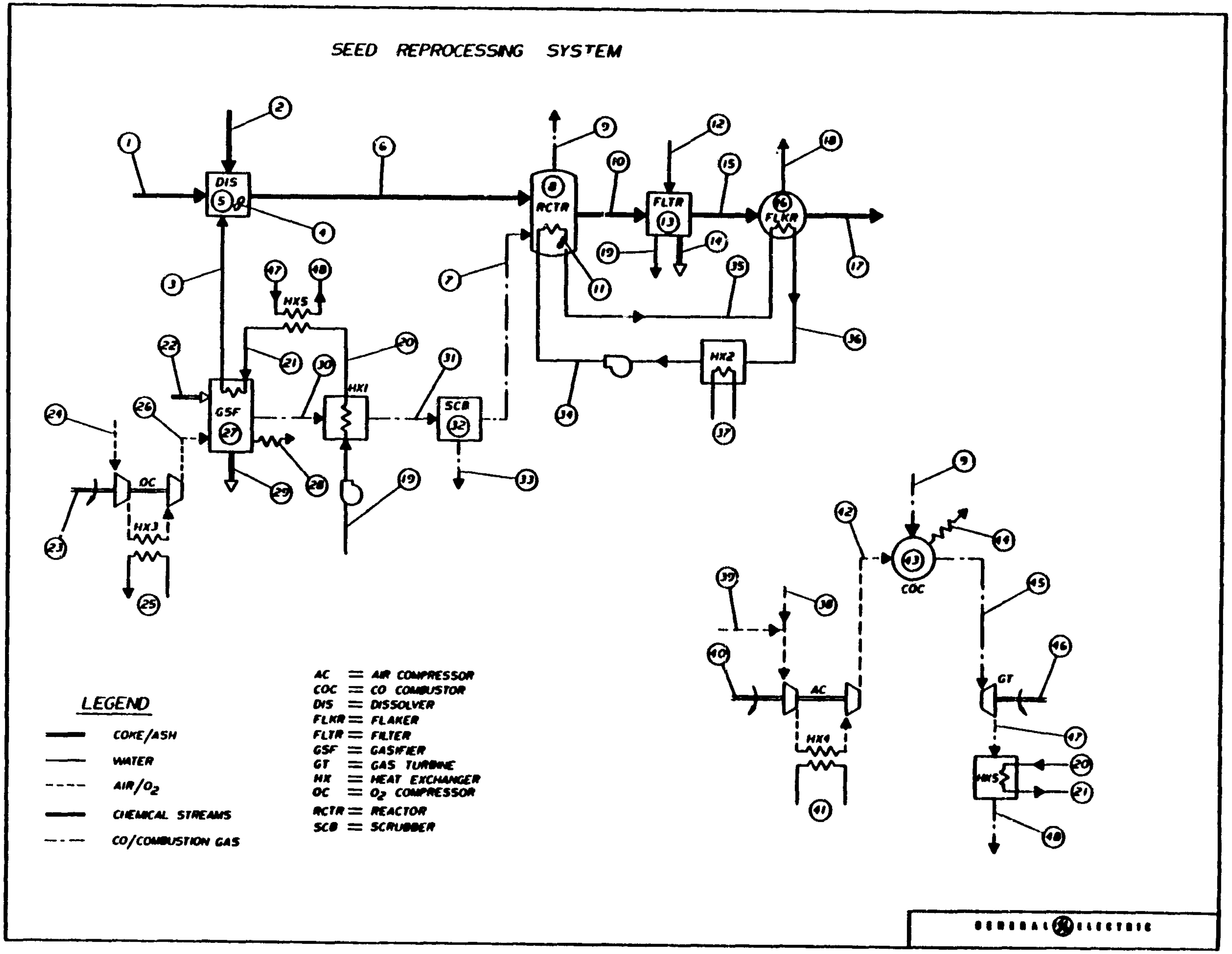


Figure 3.9-1. Diagram of the Seed Reprocessing System

Table 3.9-1. Mass and Energy Balances for Seed Reprocessing Facility

Basis: 1 kg/s ..lfur fired to combustors
95% sec. recovery

LOCATION	DESCRIPTION	T (F)	P (PSIA)	$\dot{m}$ (kg/S)	h_s (MJ/kg)	$\dot{E}_u$ (MW)	$\dot{\Delta H}_{CHEM}$ (MW)	COMMENTS
1	K_2SO_4, Reactant K_2SO_4, Excess Σ	 400	 14.7	5.1860 1.0336 6.2016	 0.172	 1.064		T = 400 F assumes K_2SO_4 cools during time from HRSR. Output (600 F) to seed reprocessing input
2	$(OH)_2$	77	14.7	2.2050	0.0	0.0		
3	H_2O to Dissolver	392	500	6.2016	0.745	4.6202		Temperature @ which reactor operates (200 C)
4	Agitator input power					0.4		Scaled from Hooker # (40-60 split between dissolver and reactor)
5	Dissolver						0.0	No reactions assumed
6	Unreacted mixture from dissolver	400	450	14.6082	0.4165	6.0844		
7	CO, Reactant CO, Excess Σ CO to reactor	 200	 450	1.6631 0.4157 2.0788	 0.0715	 0.1486		
8	Reactor						9.6321	$K_2SO_4 + Ca(OH)_2 + 2CO + 2H_2O \longrightarrow CaSO_4 \cdot 2H_2O + 2KCOOH$ Δh = 1.8573 MJ/kg K_2SO_4 reactant
9	CO, Excess	392	450	0.4157	0.1830	0.0761		
10	KCOOH $CaSO_4 \cdot 2H_2O$ K_2SO_4, Excess H_2O Σ from excess	 -400	 450	4.9897 5.1086 1.0336 5.1394 16.2713	 0.3812	 6.2033		Reactors - isothermal Remove heat of reaction with cooling coils Reacted mixture
11	Agitator input power					0.575		Scaled from Hooker # (40-60 split between dissolver and reactor)
12	H_2O to Filter	77	450	7.2638	0.0	0.0		

Table 3.9-1. Mass and Energy Balances for Seed Reprocessing Facility (Cont)

Basis: 1 kg/s sulfur fired to combustors
95% seed recovery

LOCATION	DESCRIPTION	T (F)	P (PSIA)	$\dot{m}$ (kg/S)	h_s (MJ/kg)	$\dot{E}_s$ (MW)	$\Delta \dot{H}_{CHEM}$ (MW)	COMMENTS
13	Filter						0.0	No reactions
14	$CaSO_4 \cdot 2H_2O$	77	14.7	5.1086	0.0	0.0		Filtrate
15	KCOOH			4.9897				
	K_2SO_4			1.0366				
	H_2O			6.2016				
	Σ solution	-350	400	12.2279	0.5072	6.2033		From filter
16	Flaker						0.0	No reaction; heated by water loop $\dot{Q}_{in}$ = 9.2264 MW
17	KCOOH			4.9897				
	K_2SO_4			1.0366				
	Σ Solids	301	14.7	6.0263	0.1131	0.6812		Pressure let down to 1 atm. Flaker output (H_2O and solids) @ 150 C. KCOOH contains 13.645 MW of Chem energy ($KCOOH + O_2 \rightarrow K_2CO_3 + H_2O + CO_2$, Δh = 2.7346 MJ/kg KCOOH)
18	H_2O vapor from flaker	301	14.7	6.2016	2.6656	16.5310		
19	H_2O to HX1	77	650	6.2016	0.0	0.0		
20	H_2O from HX1	260	600	6.2016	0.4280	2.6540		Heat addition from gasifier gases = 2.6540 MW
21	H_2O from HX5	360	550	6.2016	0.6568	4.0753		Heat addition from gas turbine exhaust = 1.4193 MW
22	Carbon			0.9378				
	Ash			0.1042				10% ash by weight
	Σ Coke	77	14.7	1.0420	0.0	0.0		To gasifier
23	O_2 compressor input power					0.5861		
24	O_2 to compressor	77	14.7	1.3126	0.0	0.0		Intercooled
25	Intercooler heat rejection					0.2930		HX3
26	O_2 input to gasifier	501	500	1.3126	0.2232	0.2930		

Table 3.9-1. Mass and Energy Balances for Seed Reprocessing Facility (Cont)

Basis: 1 kg/s sulfur fired to combustors
95% seed recovery

LOCATION	DESCRIPTION	T (F)	P (PSIA)	$\dot{m}$ (kg/S)	h_s (MJ/kg)	$\dot{E}_s$ (MW)	$\Delta\dot{H}_{CHEM}$ (MW)	COMMENTS
27	Gasifier						3.5582	$2.6316C + 1.3812O_2 \rightarrow 2.5CO + 0.1316CO_2$ $\Delta h = 3.7942$ MJ/kg C
28	Gasifier heat loss to ambient					0.3558		10% of reaction energy
29	Ash rejection	2500	14.7	0.1042	1.3	0.1355		
30	CO CO_2 Σ from gasifier	2060	480	2.0782 0.1720 2.2502	1.2509	2.8148		
31	CO CO_2 Σ	200	470	2.0782 0.1720 2.2502	0.0715	0.1609		Outlet gas from HX1
32	Scrubber						0.0	No reaction
33	CO_2	200	450	0.1720	0.0617	0.0106		Scrubbed from CO/CO_2 mixture
34	H_2O Loop	308	600	45.555	0.5415	19.024		
35	H_2O Loop	400	550	45.55	0.7646	29.185		10.161 MW removed frm reactor
36	H_2O Loop	300	500	45.55	0.5229	19.959		11.009 MW added to flaker
37	Heat addition for flaker					0.848		HX2
38	O_2 N_2 Σ	77	14.7	0.2374 0.7898 1.0272	0.0	0.0		Air to CO combustor
39	N_2	77	14.7	2.4230	0.0	0.0		N_2 from O_2 Plant
40	Compressor input power					1.6352		Intercooled

Table 3.9-1. Mass and Energy Balances for Seed Reprocessing Facility (Cont)

Basis: 1 kg/s sulfur fired to combustors
95% seed recovery

LOCATION	DESCRIPTION	T (F)	P (PSIA)	$\dot{m}$ (kg/S)	h_s (MJ/kg)	$\dot{E}_s$ (MW)	$\Delta\dot{H}_{CHEM}$ (MW)	COMMENTS
41	Intercooler heat rejection					0.8176		HX4
42	O_2			0.2374				
	N_2			3.2128				
	Σ	485	450	3.4502	0.237	0.8176		Oxidizer to CO combustor
43	CO combustor						5.3935	$0.5CO + 0.25O_2 \rightarrow 0.5CO_2$ per mole K_2SO_4 $\Delta h^2 = 12.9745$ MJ/kg CO
44	CO combustor heat loss					0.3144		5% of input sens + chem energy
45	CO_2			0.6531				
	N_2			3.2128				
	Σ	2500	450	3.8659	1.545	5.9728		
46	Gas turbine shaft output					3.9065		
47	Turbine exhaust	986	14.7	3.8659	0.5345	2.0663		
48	Exhaust to ambient	375	14.7	3.8659	0.1674	0.6470		

Mass balances for all PSPEC cases of interest* are shown in Table 3.9-2. These results were derived by scaling the mass balance shown in Table 3.9-1 to the potassium sulfate processing requirements for each PSPEC case of interest. It was assumed that 95% recovery of spent seed could be obtained in the HRSR system.

Energy requirements, capital costs, material costs and plant size are shown in Table 3.9-3 for each PSPEC case considered. Material costs for the cesium case are not included because a realistic price for cesium carbonate could not be obtained during the present study. Quotes were obtained from vendors only for chemically pure cesium carbonate in small quantities. Sufficient raw material in the form of pollucite and other cesium ores exist but in the absence of any significant market, industrial grade cesium salts are not presently available in quantity. Materials prices used to derive the potassium processing costs from the mass balances are given in Table 3.9-4.

Table 3.9-5 summarizes effects on overall plant efficiency and capital cost. The data are approximate but clearly indicate that, if the formate process can be implemented, the penalties for satisfying sulfur emission requirements will be quite small for an MHD/steam power plant.

3.10 STEAM PLANT PERFORMANCE

In previous large-scale MHD plant studies, the steam plant was treated either as a "black box" with a fixed efficiency or analyzed separately. For this parametric study, however, a greater degree of flexibility was required to accommodate changes in the system configuration. Subject to the 1977 consent decree (see steam plant calculation note, pg vi), a comprehensive heat balance for a modern generic steam plant was incorporated into the overall system analysis. The heat balance is calculated by means of a set of subroutines describing each of the major components of the steam plant. Internal checks are made on energy and mass balances. Water/steam properties are determined by the formulation adopted by the International Formulations Committee of the 6th International Conference on the Properties of Steam based on the 1967 ASME Steam Tables. This system of subroutines not only provides a high degree of configuration flexibility, it also assures that the bottoming cycle analysis is consistent in nomenclature and data storage formats with the topping cycle system analysis. Performance data is taken from Spencer, Cotton and Cannon[1]. Empirical curve fits were used for the expansion lines.

The general configuration of the 3500 psi/1000 F/1000 F steam turbine system is shown in Figure 3.10-1. It includes a high pressure turbine, a double flow reheat turbine and three low pressure turbines. A separate boiler feed pump turbine has been provided bu the split in total shaft power between main compressor drive and electric power generation has not been detailed.

Figure 3.10-2 shows the feedwater heating train. MHD generator cooling is done at intermediate pressure downstream of the deaerator. A low temperature economizer replaces one feedwater heater. For specific plant configurations one other of the five feedwater heaters shown also drops out.

*For Base Case 1, seed was K_2SO_4 and flue gas scrubbing was used instead of seed for sulfur capture.

Table 3.9-2. Mass Balances for the Seed Reprocessing System Required for each PSPEC Case

CASE #	COAL TYPE	COAL FLOW RATE TO MAIN COMBUSTOR (Kg/S)	SULFUR INPUT (Kg/S)	K_2SO_4 REPROCESSED (ASSUMES 95% RECOVERY IN HRSR) (Kg/S)	K_2CO_3 MAKE-UP (Kg/S)	CaO REQUIRED (Kg/S)	H_2O MAKE-UP (Kg/S)	COKE REQUIRED (Kg/S)	O_2 REQUIRED (Kg/S)	$CaSO_4$ PRODUCED (Kg/S)	FORMATE PRODUCED (17.2% K_2SO_4 IMPURITY) (Kg/S)
2.0	MR	70.36	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
2.0S	MR	70.36	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
2.0A	MR	42.58	0.445	2.301	0.176	0.743	3.234	0.464	0.584	1.799	2.683
2.0B	MR	61.11	0.585	3.026	0.0232	0.977	4.253	0.610	0.769	2.366	3.529
2.1	I6	63.98	2.270	11.74	0.0900	3.792	16.501	2.367	2.982	9.178	13.690
2.2	MR	70.36	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
2.2A	MR	70.36	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
2.4	MR	70.36	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
2.4A	MR	70.36	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
2.5 (Cesium)	MR	70.36	0.735	10.722	0.3734	1.228	5.342	0.766	0.965	2.972	9.325
2.6	MR	70.36	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
2.7	MR	70.36	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
2.10	MR	37.69	0.394	2.038	0.0156	0.658	2.864	0.411	0.518	1.593	2.376
2.11	MR	50.25	0.525	2.715	0.0208	0.877	3.816	0.547	0.690	2.123	3.166
2.11A	MR	50.25	0.525	2.715	0.0208	0.877	3.816	0.547	0.690	2.123	3.166
2.12	MR	70.36	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
2.16	MR	70.36	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
2.16A	MR	70.36	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
2.16B	MR	70.36	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
2.17	MR	70.35	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
2.18	MR	70.36	0.735	3.801	0.0291	1.228	5.342	0.766	0.965	2.972	4.432
3.0	MR	98.32	1.027	5.311	0.0407	1.716	7.464	1.070	1.348	4.153	6.193
3.1	I6	89.34	3.170	16.394	0.1255	5.296	23.04	3.304	4.162	12.82	19.12
3.2	MR	98.32	1.027	5.311	0.0407	1.716	7.464	1.070	1.348	4.153	6.193
3.4	MR	100.00	1.045	5.404	0.0414	1.746	7.595	1.089	1.372	4.225	6.301
3.5	MR	98.32	1.027	5.311	0.0407	1.716	7.464	1.070	1.348	4.153	6.193

Table 3.9-3. Energy Requirements, Capital Cost, Operating Cost and Plant Size for each PSPEC Case

CASE #	THERMAL INPUT OF COKE (E_{TH} = 29.5 MJ/Kg) (MWt)	OTHER THERMAL REQUIREMENTS (MWt)	SHAFT POWER REQUIRED (MWe)	THERMAL CREDIT FOR FORMATE (MWt)	TOTAL THERMAL PENALTY (MWt)	CAPITAL COST MILLIONS	LABOR AND MATERIALS MILLS/KWH	PLANT SIZE SQ. METERS
2.0	22.618	0.624	0.435	10.036	14.20	10.99	0.515	7038
2.0S	22.618	0.624	0.435	10.036	14.21	10.99	0.519	7038
2.0A	13.701	0.378	0.263	6.075	8.62	7.64	0.382	4261
2.0B	18.011	0.497	0.346	7.990	11.32	9.27	0.471	5603
2.1	69.890	1.926	1.342	30.997	43.84	28.50	1.350	21,738
2.2	22.618	0.624	0.435	10.036	14.19	10.99	0.555	7038
2.2A	22.618	0.624	0.435	10.036	14.21	10.99	0.519	7038
2.4	22.618	0.624	0.435	10.036	14.21	10.99	0.521	7038
2.4A	22.618	0.624	0.435	10.036	14.21	10.99	0.516	7038
2.5 (CESIUM)	22.618	0.624	0.435	9.44	14.77	10.99		7038
2.6	22.618	0.624	0.435	10.036	14.22	10.99	0.516	7038
2.7	22.618	0.624	0.435	10.936	14.17	10.99	0.501	7038
2.10	12.136	0.334	0.233	5.381	7.64	7.03	0.627	3774
2.11	16.151	0.445	0.310	7.168	10.15	8.58	0.565	5027
2.11A	16.151	0.445	0.310	7.168	10.16	8.58	0.569	5027
2.12	22.618	0.624	0.435	10.036	14.22	10.99	0.530	7038
2.16	22.618	0.624	0.435	10.036	14.16	10.99	0.538	7038
2.16A	22.618	0.624	0.435	10.036	14.18	10.99	0.549	7038
2.16B	22.618	0.624	0.435	10.036	14.18	10.99	0.508	7038
2.17	22.618	0.624	0.435	10.036	14.23	10.99	0.505	7038
2.18	22.618	0.624	0.435	10.036	14.19	10.99	0.548	7038
3.0	31.57	0.872	0.608	14.02	19.83	14.30	0.781	9834
3.1	97.47	2.691	1.876	43.29	61.16	38.22	2.126	30,355
3.2	31.57	0.872	0.608	14.02	19.83	14.30	0.781	9834
3.4	32.13	0.887	0.618	14.27	20.19	14.50	0.785	10,006
3.5	31.57	0.872	0.608	14.02	19.79	14.30	0.760	9834

Table 3.9-4. Materials Prices used to Derive Potassium Processing Costs

K_2CO_3	484 $/T	0.535 $/KG
CaO	35 $/T	0.0386 $/KG
COKE	90 $/T	0.0992 $/KG
$CaSO_4$ (DISPOSAL)	0.90 $/T	0.001 $/KG
LABOR: 7 MEN, 14,000 /YR		

Table 3.9-5. Summary of Seed Reprocessing Requirements for Example from Base Case 3

REQUIREMENTS	MONTANA ROSEBUD	ILLINOIS #6
THERMAL INPUT OF COKE (MWT)	22.6	69.9
OTHER THERMAL REQUIREMENTS (MWT)	0.6	1.9
SHAFT POWER (MWE)	0.4	1.3
CREDIT FOR FORMATE (MWT)	10.0	31.0
ENERGY DEBIT (MWT)	14.2	43.8
$\sim\Delta\eta$	0.3	0.9
CAPITAL COSTS* ($ MILLIONS)	11.0	28.5
$\sim\Delta$ $/KW	20	51

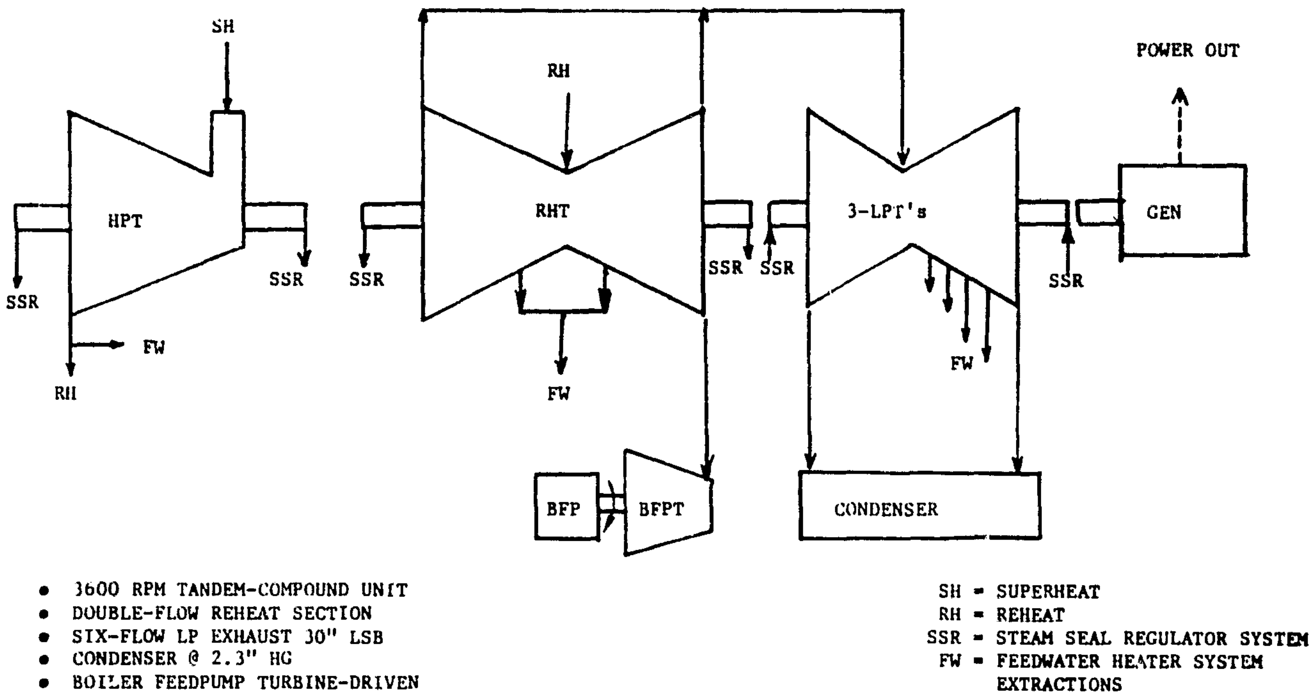


Figure 3.10-1. Steam Turbine/Generator Configuration

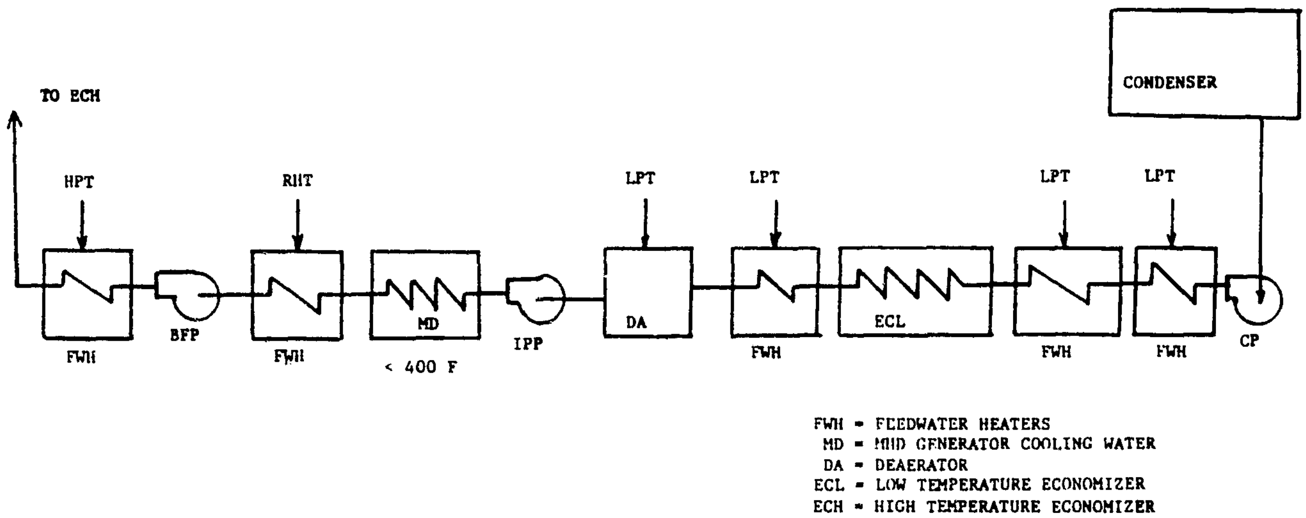


Figure 3.10-2. Feedwater Heater System

SECTION 4

COSTING

SECTION 4

COSTING

4.1 GENERAL GROUND RULES

For consistency and ease of comparison between past and future studies, the cost estimate formats, guidelines and the modified FPC code of accounts used for the ETF conceptual design study[1] were used in this study. The code of accounts is included as Appendix J. Unless otherwise noted, all costs given in this report are expressed in terms of mid-1978 dollars. Those costs which were obtained by scaling appropriate values from past studies[2], such as ECAS I and II or the ETF designs, were escalated to mid-1978 dollars by applying a fixed rate of 8% per year to the estimated cost. Numerical values of constants used for cost calculations are included in an example case, Appendix G.

4.2 CAPITAL COST

Conceptual capital cost estimates were prepared for the engineering and construction of the reference design case and all parametric variations considered for each of the three Base Case MHD Power Plants. The capital costs of the major MHD system components for all reference cases and parametric variations were estimated by the General Electric Company. For the three reference cases, the remaining capital costs were estimated by Bechtel National Inc. The Bechtel estimates were based on the conceptual design information supplied by GE and costing methods extrapolating from Bechtel's current cost data. On the basis of the Bechtel estimates, the three reference cases were updated and the capital cost for the remaining parametric cases were obtained by applying appropriate scaling factors to account for the differences between the various systems.

The capital cost estimates are composed of field construction costs, engineering services, contingency and escalation and interest during construction. The largest category, field construction costs, includes the direct cost of permanent plant equipment and indirect cost of temporary construction materials. A complete listing of capital costs for all cases is included as Appendix H.

4.2.1 MAJOR EQUIPMENT

In general, all major equipment costs quoted by this study are direct costs for the components delivered to the site. These costs include the costs of auxiliary components, instrumentation and control, but do not include installation or field erection unless specifically noted for a particular component.

The cost of the major MHD system equipment was estimated by GE. The MHD channel costs were scaled up from the GE ETF Study[3] and escalated to mid-1978 dollars. Costs for the HRSR system components were estimated on a dollars per pound basis, after first determining

the furnace size as described in Section 3.7.3 and selecting the proper construction materials. Costing of the MHD magnet and dewar is discussed in Section 3.4. The basis for estimating the costs of the MHD combustors, HTAH assemblies and HTAH gasifiers is provided in Appendix F.

The following major equipment costs were supplied by Bechtel: coal handling equipment, seed injection equipment, main compressor, and the steam turbine generator set. A listing of the major equipment costs for all cases is shown in Tables 4.2-1, 4.2-2 and 4.2-3.

4.2.2 BALANCE OF PLANT EQUIPMENT AND MATERIALS

The costs of balance of plant equipment and materials were estimated based on previous work done by Bechtel in the ECAS II Study[4] and escalated to mid-1978 dollars. The balance of plant costs were developed by adjusting the ECAS data to account for differences in the size or capacity of major items or systems.

4.2.3 INSTALLATION COSTS

The labor required to install major equipment was estimated based on equipment size or capacity using current Bechtel engineering data. The installation labor required for the balance of plant equipment and material was estimated based on the ECAS work. Installation cost was computed using the Middletown mid-1978 composite wage rate of $14.30 per manhour, which represents the average United States Labor rate.

4.2.4 INDIRECT COSTS

The indirect field costs are those items of construction cost that cannot be ascribed directly to portions of the facility and are thus accounted separately. The items covered under this category include: temporary construction facilities, construction equipment and supplies, miscellaneous construction services, preliminary checkout and acceptance testing, and project insurance. These indirect costs were estimated, based on Bechtel's experience in constructing fossil-fired plants, at 75% of the direct installation labor costs.

4.2.5 ENGINEERING SERVICES

The engineering services include engineering costs, other home office costs and fee. These costs were estimated at 15% of the combined total of balance of plant, installation and indirect costs. This was the same method used for estimating the engineering services cost in the ECAS II study[5].

4.2.6 CONTINGENCY

A contingency cost is included in the estimate as an allowance for the uncertainty that exists within the conceptual design in quantity, pricing or productivity, and is under the control of

Table 4.2-1. Major Equipment Costs Mid-78 Dollars x 10^6

	COMPONENTS	Case 1.0	Case 1.1	Case 1.2	Case 1.3	Case 1.4	Case 1.4a
312	BOILER PLANT EQUIP.						
312.3	Radiant Furnace	13.8	13.8	14.2	13.4	14.0	13.7
312.4	Final Oxidation Furnace	29.5	29.5	30.4	28.6	30.0	29.4
	Reheaters	11.2	11.1	11.5	10.8	11.3	11.1
	Initial Superheater	1.3	1.3	1.4	1.3	1.3	1.3
	Economizers	9.8	9.8	10.1	9.5	10.0	9.8
312.5	Electrostatic Precipitator	6.3	6.0	7.2	6.4	6.4	6.4
	Sulfur Removal System	13.9	13.9	13.9	13.9	13.9	13.9
	Subtotal	85.8	85.4	88.7	83.9	86.9	85.6
314.1	STEAM TURBINE GENERATOR	27.9	27.3	29.2	27.0	27.9	28.6
317	MHD TOPPING CYCLE						
317.1	Main Combustor	10.7	9.8	14.3	9.9	6.8	5.8
317.2	MHD Channel	18.3	18.1	19.8	18.2	18.3	18.5
	Diffuser	6.2	6.3	6.7	6.3	6.3	6.2
317.3	Magnet/Dewar	116.0	116.0	116.0	200.0	116.0	116.0
317.5	HTAH Assembly	131.6	131.6	159.8	131.6	131.6	131.6
	HTAH Gasifier	10.5	9.0	12.6	10.5	10.5	9.0
	Air Heaters	2.5	3.3	3.2	4.0	3.3	3.3
	Main Compressor/Turbine	11.7	11.4	11.4	11.9	11.5	11.1
	Preheat Compress./Turbine	-	-	-	-	-	-
317.4	Electrical Inversion Equip.	43.2	45.7	39.0	46.4	43.8	41.9
*312.1	Coal Handling Equipment	25.4	22.4	25.8	25.4	25.4	25.4
317.6	Seed Handling Equipment	1.4	1.4	1.4	1.4	1.4	1.4
	Seed Reprocessing Equip.	-	-	-	-	-	-
	Subtotal	377.5	375.5	410.0	465.6	374.9	370.2
	TOTAL, EXCLUDING O_2 PLANT	491.2	488.4	527.9	576.5	489.7	484.4
317.9	OXYGEN PLANT	33.0	33.0	-	33.0	33.0	33.0

*Included in Account 317 because BOP estimate lumped coal handling and coal injection

Table 4.2-2. Major Equipment Costs Mid-78 Dollars x 10^6

	COMPONENTS	Case 2.0	Case 2.0s	Case 2.0a	Case 2.0b	Case 2.1	Case 2.2	Case 2.2a	Case 2.4	Case 2.4a	Case 2.5	Case 2.6	Case 2.7	Case 2.10	Case 2.11	Case 2.11a	Case 2.12	Case 2.15	Case 2.16	Case 2.16a	Case 2.16b	Case 2.17	Case 2.18
312	BOILER PLANT EQUIP.																						
312.3	Radiant Furnace	15.4	15.7	14.3	14.0	15.8	14.5	15.6	15.5	15.5	14.9	15.8	14.9	8.7	11.3	11.5	15.7	15.4	14.0	14.3	15.0	15.6	14.5
312.4	Final Oxidation Furnace	33.0	33.5	30.7	30.1	33.8	31.1	33.4	33.3	33.3	31.9	33.9	31.9	18.6	24.2	24.6	33.5	33.0	30.0	30.7	32.0	33.4	31.1
312.4	Reheaters	12.5	12.7	11.6	11.4	12.8	11.7	12.6	12.6	12.6	12.1	12.8	12.0	7.0	9.1	9.3	12.7	12.5	11.3	11.6	12.1	12.6	11.7
312.4	Initial Superheater	1.5	1.5	1.4	1.3	1.5	1.4	1.5	1.5	1.5	1.4	1.5	1.4	0.8	1.1	1.1	1.5	1.5	1.3	1.4	1.[illegible]	1.5	1.4
312.4	Economizers	13.1	13.4	10.2	10.0	12.5	10.4	13.2	13.2	13.2	12.7	13.5	12.7	7.4	9.6	9.7	11.2	13.1	10.0	10.2	12.3	13.3	10.4
312.5	Electrostatic Precipitator	4.5	4.5	4.8	-	4.5	4.5	4.5	4.5	4.5	4.5	4.5	4.5	2.7	3.4	3.4	4.5	4.5	4.5	4.5	4.5	4.5	4.5
312.5	Sulfur Removal System	3.3	3.3	2.6	1.9	3.2	2.6	3.3	3.2	3.3	3.2	3.4	3.3	1.8	2.4	2.4	3.2	3.3	2.6	2.6	3.[illegible]	-	2.6
	Subtotal	83.3	84.7	75.6	68.7	84.1	76.2	84.1	83.8	83.9	80.7	85.4	80.7	47.0	61.1	62.0	82.3	83.3	73.7	75.3	81.[illegible]	80.9	76.[illegible]
314.1	STEAM TURBINE GENERATOR	28.5	29.3	26.2	26.1	28.7	26.0	28.8	28.0	29.0	27.1	30.2	27.5	18.6	22.6	23.3	27.6	28.5	24.5	25.7	27.[illegible]	29.3	25.8
317	MHD TOPPING CYCLE																						
317.1	Main Combustor	6.7	6.7	1.0	33.0	6.7	14.5	14.5	11.2	11.2	5.8	9.0	6.3	4.7	5.6	5.6	16.1	13.4	5.5	6.7	5.[illegible]	6.7	6.7
317.2	MHD Channel	19.4	19.6	20.3	20.5	19.3	19.5	19.5	15.4	15.4	19.2	22.0	19.4	8.7	13.3	13.5	19.4	38.8	19.2	19.4	19.[illegible]	19.4	19.4
317.2	Diffuser	6.7	6.6	6.7	6.9	6.6	6.6	6.6	6.3	6.3	6.7	8.3	6.6	3.4	4.7	4.6	6.7	13.4	6.8	6.7	6.[illegible]	6.7	6.7
317.3	Magnet/Dewar	116.0	116.0	116.0	116.0	116.0	116.0	116.0	116.0	116.0	116.0	148.4	296.3	67.7	91.5	91.5	116.0	226.2	116.0	116.0	116.[illegible]	116.0	116.0
317.5	HTAH Assembly	136.2	136.2	142.7	109.0	125.4	89.0	136.2	126.2	126.2	114.0	190.5	125.3	76.0	100.1	100.1	181.5	143.0	66.8	66.8	136.[illegible]	126.3	136.2
317.5	HTAH Gasifier	5.9	5.9	34.0	24.0	4.7	5.1	5.1	5.6	5.6	4.9	7.6	5.6	3.5	4.7	4.7	18.0	6.2	3.7	3.7	5.[illegible]	3.0	3.7
317.5	Air Heaters	-	-	4.7	5.6	-	9.1	-	-	-	-	-	-	-	-	-	2.9	-	8.6	9.0	-	-	3.8
317.5	Main Compressor/Turbine	12.0	11.4	11.6	11.6	11.9	12.2	11.8	12.7	11.6	12.5	10.6	12.0	7.3	9.2	8.7	11.9	12.0	13.0	12.3	12.	12.0	12.0
317.5	Preheat Compress./Turbine	12.5	12.7	8.3	6.7	12.3	11.4	12.5	12.4	12.6	12.4	12.8	12.6	8.4	10.2	10.3	-	12.5	11.3	10.7	12.[illegible]	12.0	11.2
317.4	Electrical Inversion Equip.	43.4	41.1	40.0	39.9	43.8	42.1	42.1	42.5	42.5	47.6	39.9	47.8	22.7	31.7	29.9	43.4	43.4	47.2	43.4	47.	43.4	38.6
*312.1	Coal Handling Equipment	24.2	24.3	24.0	24.0	22.3	22.6	24.2	24.0	24.3	24.1	24.5	24.2	14.8	18.6	18.6	23.9	38.7	22.6	-2.6	24.[illegible]	25.1	22.8
317.6	Seed Handling Equipment	1.3	1.3	1.3	1.3	1.3	1.3	1.3	1.3	1.3	1.3	1.3	1.3	0.8	1.0	1.0	1.3	1.3	1.3	1.3	1.[illegible]	1.3	1.3
317.6	Seed Reprocessing Equip.	11.0	11.0	7.6	9.3	23.5	11.0	11.0	11.0	11.0	11.0	11.0	11.0	7.0	8.6	8.6	11.0	11.0	11.0	11.0	11.[illegible]	11.0	11.0
	Subtotal	395.3	392.8	418.2	407.8	398.8	360.4	400.8	384.6	383.9	375.5	485.9	568.4	225.0	299.2	297.1	442.1	559.9	333.0	329.6	498.[illegible]	381.9	389.4
	TOTAL, EXCLUDING O_2 PLANT	507.1	506.8	520.0	502.6	511.6	462.6	513.7	496.4	496.8	483.3	601.5	676.6	290.6	382.9	382.4	552.0	671.7	431.2	430.6	506.	492.1	491.4
317.9	OXYGEN PLANT	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

*Included in Account 317 because BOP estimate lumped coal handling and coal injection

Table 4.2-3. Major Equipment Costs Mid -78 Dollars x 10^6

	COMPONENTS	Case 3.0	Case 3.1	Case 3.2	Case 3.4	Case 3.5
312	BOILER PLANT EQUIP.					
312.3	Radiant Furnace	14.7	14.2	14.5	15.1	13.9
312.4	Final Oxidation Furnace	31.5	30.6	31.0	32.4	29.8
	Reheaters	11.9	11.5	11.7	12.2	11.3
	Initial Superheater	1.4	1.4	1.4	1.4	1.3
	Economizers	10.5	10.2	10.3	10.8	9.9
312.5	Electrostatic Precipitator	3.9	3.9	4.0	4.1	4.0
	Sulfur Removal System	-	-	-	-	-
	Subtotal	73.9	71.8	72.9	76.0	70.2
314.1	STEAM TURBINE GENERATOR	28.2	26.8	27.4	28.9	25.8
317	MHD TOPPING CYCLE					
317.1	Main Combustor	8.9	7.4	12.8	9.0	6.4
	MHD Channel	17.5	17.1	17.3	17.6	17.1
317.2	Diffuser	5.6	5.7	5.7	5.6	5.8
317.3	Magnet/Dewar	107.0	107.0	107.0	107.0	273.3
317.5	HTAH Assembly	-	-	-	-	-
	HTAH Gasifier	-	-	-	-	-
	Air Heaters	4.3	4.1	4.2	3.3	3.8
	Main Compressor/Turbine	9.7	10.6	10.4	9.8	11.1
	Preheat Compress./Turbine	-	-	-	-	-
317.4	Electrical Inversion Equip.	37.0	40.5	38.6	36.4	43.4
*312.2	Coal Handling Equipment	21.9	20.1	21.7	22.0	21.7
317.6	Seed Handling Equipment	1.3	1.3	1.3	1.3	1.3
	Seed Reprocessing Equip.	11.0	38.2	14.3	14.3	14.3
	Subtotal	224.2	252.0	233.3	331.2	398.2
	TOTAL, EXCLUDING O_2 PLANT	326.3	350.6	333.6	340.2	494.2
317.9	OXYGEN PLANT	85.6	85.4	85.6	87.0	85.6

*Included in Account 317 because BOP estimate lumped coal handling and coal injection

the constructor while still being within the scope of the project as defined. A contingency of 20% of the field construction cost plus the appropriate percentage of the engineering services cost was applied to the MHD portions of the system. A contingency of 10% was applied to the remaining conventional power plant portions of the system. The higher contingency applied to the MHD system equipment is attributable to the greater uncertainty that exists in its design and construction when compared with more conventional power plant equipment.

4.2.7 ESCALATION AND INTEREST DURING CONSTRUCTION

The cost of escalation and interest during construction in mid-1978 dollars is included in the capital cost estimate. A 6-1/2 year engineering and construction period was assumed for each plant. Based on NASA guidelines, a 6-1/2% future escalation rate and a 10% interest rate was also assumed along with an S-shaped cash flow distribution during the construction period. These assumptions are consistent with those used in the ECAS II study.

4.2.8 OXYGEN PLANT COSTS

For the system designs requiring oxygen enrichment in Base Case 1 and Base Case 3, the cost of the on-site oxygen plant has been included in the capital cost estimates. The prices quoted for these oxygen plants were based upon mid-1978 vendor price data, and reflect the turnkey price for the appropriately sized oxygen plant. Consequently, the oxygen plant was included as a separate item in the estimate, and no balance of plant, labor, engineering services, or contingency costs were applied to the oxygen plant costs.

The capital cost summaries for all the cases considered in PSPEC are listed in Appendix H. Capital cost comparisons between the three reference cases and the key parametric variations of these designs are listed in Tables 4.2-4 through 4.2-7.

4.3 COST OF ELECTRICITY

The cost of electricity in mills per kilowatt-hours, was calculated for each case. This cost of electricity is a life cycle cost in the sense that it is the average cost of the energy produced during the plant lifetime. All costs were based on a thirty-year plant life and a 65 percent plant capacity factor. In calculating these costs, the year 2000 was assumed to be the date for the start of commercial operation with a 6-1/2% general inflation rate assumed throughout the life of the plant.

The overall procedure used in computing the cost of electricity is described below. First, all the costs for capital investment, fuel and operation and maintenance expenses are estimated in mid-1978 dollars. These costs are then inflated at their appropriate rates to the date for the start of commercial operation. The cost of electricity is then levelized over the life of the plant using a present worth averaging technique which expresses the cost as a series of equal cash payments made of the life of the plant. Finally, these costs are expressed in terms of mid-1978 dollars by deescalating at the same rate as the general inflation rate.

Table 4.2-4. Capital Cost Comparison (Mid-78 Dollars x 10^6) References Cases

CASE NO.	1.0	2.0	3.0
BASE CASE SPECIFICATION	ATM HTAH	PRESS HTAH	O_2
CAPITAL COSTS: (MILLION $)			
FIELD CONSTRUCTION COST	896.5	873.0	682.0
ENGINEERING SERVICES	60.8	54.9	53.4
CONTINGENCY	150.3	147.1	106.9
OXYGEN PLANT	33.0	NA	85.6
ESCALATION AND INTEREST DURING CONSTRUCTION	131.1	123.6	106.7
TOTAL CAPITAL COST	1271.7	1198.5	1034.6
PLANT OUTPUT (MWe)	1189.3	1257.4	1089.3
PLANT CAPITAL COST ($/kWe)	1069.1	953.2	949.8

Table 4.2-5. Capital Cost Comparison (Mid-78 Dollars x 10^6) Plant Size Variations

CASE NO.	2.0	2.11	2.10
THERMAL POWER TO MHD COMBUSTOR	2800 MWt	2000 MWt	1500 MWt
CAPITAL COST: (MILLION $)			
FIELD CONSTRUCTION COST	873.0	663.4	517.5
ENGINEERING SERVICES	54.9	42.1	34.0
CONTINGENCY	147.1	111.7	86.6
ESCALATION AND INTEREST DURING CONSTRUCTION	123.6	94.0	73.4
TOTAL CAPITAL COST	1198.6	911.2	711.5
PLANT OUTPUT (MWe)	1257.4	887.2	654.5
PLANT CAPITAL COST ($/kWe)	953.2	1027.1	1087.1

Table 4.2-6. Capital Cost Comparison (Mid-78 Dollars x 10^6) Combustion Variations

CASE NO.	2.0	2.0a	2.1	2.16
PRIMARY CHANGE FROM REFERENCE CASE	REF.	S^3PMB + COAL	ILL. #6	4 kV/m + HOT BOTTOM HTAH
CAPITAL COSTS: (MILLION $)				
FIELD CONSTRUCTION COST	873.0	878.4	873.6	771.6
ENGINEERING SERVICES	54.9	53.8	54.3	51.1
CONTINGENCY	147.1	150.5	146.9	129.0
ESCALATION AND INTEREST DURING CONSTRUCTION	123.6	124.5	123.6	109.5
TOTAL CAPITAL COST	1198.6	1207.2	1198.4	1061.2
PLANT OUTPUT (MWe)	1257.4	1165.4	1272.0	1202.9
PLANT CAPITAL COST ($/kWe)	953.2	1035.9	942.1	882.2

Table 4.2-7. Capital Cost Comparison (Mid-78 Dollars x 10^6) MHD Flow Train Variations

CASE NO.	2.0	2.7	2.15
PRIMARY CHANGE FROM REFERENCE CASE	REF.	(8-7) T MAGNET	DUAL FLOW TRAIN
CAPITAL COSTS: (MILLION $)			
FIELD CONSTRUCTION COST	873.0	1044.1	1135.4
ENGINEERING SERVICES	54.9	55.1	69.6
CONTINGENCY	147.1	181.9	198.6
ESCALATION AND INTEREST DURING CONSTRUCTION	123.6	147.3	161.4
TOTAL CAPITAL COST	1198.6	1428.4	1565.0
PLANT OUTPUT (MWe)	1257.4	1293.2	1257.4
PLANT CAPITAL COST ($/kWe)	953.2	1104.5	1244.6

The levelization procedure and cost equations used in this study are described in Chapter 3 of the EPRI report entitled "Comparative Study and Evaluation of Advanced Cycle Systems."[6] Using a 30 year plant life, a 6-1/2% annual inflation rate and a weighted cost of capital equal to 10% per year, the calculated value of the cost levelization factor is 1.882. This was the value used in levelizing the fuel and operation and maintenance costs for each of the PSPEC cases.

4.3.1 CAPITAL COSTS

The capital cost contribution to the cost of electricity was determined according to NASA guidelines by applying a fixed charge rate of 18%/year to the total capital cost.

4.3.2 FUEL COSTS

The fuel cost portion of the cost of electricity was determined for each case using an average mid-1978 coal price of $1.05/MBTU. This price was assumed to escalate at a rate equal to the general inflation rate of 6-1/2% per year throughout the construction period, and continue on at this rate throughout the life of the plant. Using these assumptions the resulting fuel costs were levelized over the life of the plant. These levelized costs were then de-escalated to be expressed in mid-1978 dollars.

To determine the sensitivity of the cost of electricity to increases in fuel costs, a range of different coal prices and various annual inflation rates were considered over the life of the plant. The fuel costs varied from a low value of $.50/MBTU in mid-1978 to a high value of $1.35/MBTU. The range of coal inflation rates considered extends up to the case where coal prices increase at a rate of 5% higher than the general inflation rate (11-1/2%). The effects of these coal price variations on the levelized fuel cost of electricity are shown in Table 4.3-1 for two different plant operating efficiencies.

4.3.3 OPERATION AND MAINTENANCE COSTS

The operation and maintenance costs included in the cost of electricity were estimated based on previous work done by General Electric in the ECAS II study.

4.3.3.1 Maintenance Cost

The maintenance cost for conventional plant components were developed by adjusting the ECAS data to account for differences in plant sizes and escalating to mid-1978 dollars. For the MHD cycle advanced technology components, estimates of the minimum expected service lives were reevaluated, and selected maintenance methods were reviewed. The ECAS data was then adjusted to reflect changes and escalated to mid-1978 dollars. For the MHD generator, a minimum service life of 10,000 hours was assumed.

Table 4.3-1. Impact of Inflation on Fuel COE Levelized Costs, Assuming Startup in Year 2000 (Mid-78 Dollars)

		MILLS/KW HR	
FUEL COST RISE	POWER PLANT EFFICIENCY	LOW ($.50/MBTU)	HIGH ($1.35/MBTU)
INFLATION + 0%	36% 45% DIFFERENCE	8.9 7.1 1.8	24.1 19.3 4.8
INFLATION + 2.5% TO YEAR 2000 INFLATION + 0% AFTER YEAR 2000	36% 45% DIFFERENCE	14.9 11.9 3.0	40.1 32.1 8.0
INFLATION + 2.5%	36% 45% DIFFERENCE	20.1 16.1 4.0	54.2 43.4 10.8
INFLATION +5% TO YEAR 2000 INFLATION + 0% AFTER YEAR 2000	36% 45% DIFFERENCE	24.5 19.6 4.9	66.1 52.9 13.2
INFLATION +5%	36% 45% DIFFERENCE	46.1 36.9 9.2	124.5 99.6 24.9

4.3.3.2 Operating Labor Costs

In calculating the operational labor costs, the same number of plant operating personnel required for the open cycle MHD plant in the ECAS II study[7] was also assumed for the plants considered in this analysis. With the number of employees for each case estimated in this manner, the average annual salary plus fringe benefits and overhead specified in the ECAS II study was escalated at 8%/year to arrive at the total operating labor cost expressed in mid-1978 dollars.

4.3.3.3 Operating Consumables and Supplies

The operating consumables cost for the conventional plant equipment were scaled from ECAS II numbers and escalated to mid-1978 dollars. The seed makeup requirements and costs for the seed reprocessing plants were calculated using the recommended methods and cost data supplied by the Hooker Chemical Company[8]. These costs were then added to the conventional plant equipment costs to determine the total cost of operating consumables for each plant.

4.4 DISCUSSION OF RESULTS

Cost and efficiency for the three references cases are summarized in Table 4.4-1 and complete tabulations of capital and operating costs through COE for all cases are included as Tables 4.4-2 through 4.4-4. Breakdowns of capital cost by major account are included as Appendix H.

Table 4.4-1. COE and Efficiency for References Cases

CASE	1.0	2.0	3.0
BASE CASE	HTAH WITH ATMOSPHERIC PRESSURE GASIFIER FOR REHEAT	HTAH WITH PRESSURIZED REHEAT	Air + 40% O_2, NO HTAH
CAPITAL COST ($/kWe)	1069.1	953.2	949.8
COE (Mills/KwHr)	55.8	52.05	52.88
EFFICIENCY	41.4	43.4	42.9

Of the reference cases, the O_2 enriched case, Base Case 3, is intermediate between the two cases involving an indirectly-fired HTAH system*. The differences in COE are relatively small however, and the O_2 enriched approach has an advantage in that the equipment for O_2 production requires no development whereas the HTAH subsystem represents some extrapolation from present steel industry and high temperature test facility practices.

The perturbations from the reference case in Base Case 1 led to relatively minor variations (see Table 4.4-2). The apparent decrease in cost with the use of Illinois #6 coal does not reflect the difference in sulfur removal cost, since sufficient detail on the dry scrubber system was not available at the time the tables were prepared**. As previously noted, the power required for O_2 enrichment in conjunction with an HTAH system resulted in a net decrease in system output (compare Cases 1.0 and 1.2) but the reduced size of the HTAH

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* In costing the Base Case 3 magnet the reduced mass flow rate associated with O_2 enrichment was not accounted for initially. Reducing exit warm bore from 5.4 m to 5.0 m results in a saving of about $9 million in major equipment cost. A correction to COE has been made for all Base Case 3 cases.

** Updated cost estimates for the FGD system for Base Case 1 indicate that additions should be made to both capital and operating COE. For Case 1.0 (Montana Rosebud Coal) add 0.7 mills/KWhr to capital and 3.6 mills/KWhr to levelized operating cost. For Case 1.1 (Illinois #6 coal) the comparable figures are 1.2 and 4.5 mills/KWhr, respectively.

Table 4.4-2. Base Case 1, Cost Summary Through COE
(Mid-78 Dollars, Fuel and O&M Levelized)

CASE NO.	1.0	1.1	1.2	1.3	1.4	1.4a
PRIMARY CHANGE FROM REFERENCE CASE*	Ref.	I6	Air	7T	Single Stage	Slag Rejection
CAPITAL COSTS: ($ X 10^6)						
MAJOR EQUIPMENT COST	491.2	488.2	527.9	576.5	489.7	484.4
BOP MATERIAL COST	164.4	160.4	170.8	165.2	164.4	164.2
INSTALLATION COST	137.6	134.7	145.2	138.3	137.8	137.6
INDIRECT FIELD COST	103.3	101.0	108.9	103.7	103.2	103.1
FIELD CONSTRUCTION COST	896.5	884.3	952.8	983.7	895.1	889.3
ENGINEERING SERVICES	60.8	59.4	63.7	61.1	60.8	60.7
CONTINGENCY	150.3	147.9	161.4	168.2	149.9	148.8
OXYGEN PLANT	33.0	33.0	NA	33.0	33.0	33.0
ESCALATION AND INTEREST DURING CONSTRUCTION	131.1	129.3	135.4	143.3	130.9	130.1
TOTAL CAPITAL COST	1271.7	1253.9	1313.3	1389.3	1269.7	1261.9
PLANT OUTPUT (MWe)	1189.5	1214.1	1203.2	1212.3	1199.5	1192.0
CAPITAL COST ($/kWe)	1069.1	1032.8	1091.5	1146.0	1058.5	1058.6
COST OF ELECTRICITY: (Mills/KwHr)*						
CAPITAL COST	33.80	32.65	34.50	36.23	33.46	33.47
FUEL COST	16.29	15.72	16.16	15.98	16.15	16.25
OPERATION & MAINTENANCE	5.49	5.42	5.59	5.43	5.47	5.50
TOTAL COE	55.58	53.79	56.25	57.64	55.08	55.22
OVERALL PLANT EFFICIENCY (%)	41.41	42.90	41.73	42.21	41.76	41.50

*See Table 1.5-2 for details of case variations

Table 4.4-3. Base Case 2 Cost Summary Through COE
(Mid-1978 Dollars, Fuel and O&M Levelized)

CASE NO.	2.0	2.0s	2.0a	2.0b	2.1	2.2	2.2a
PRIMARY CHANGE FROM REFERENCE CASE*	Ref.	Slag Rejection	S^3PMB + Coal	S^3PMB + SPMB	I6	2-Stage Cyc. Hot Bottom HTAH	2-Stage Cyclone
CAPITAL COSTS: ($ X 10^6)							
MAJOR EQUIPMENT COST	507.1	506.8	520.0	502.6	511.6	462.6	513.7
BOP MATERIAL COST	152.7	153.1	147.9	152.8	151.4	143.4	152.8
INSTALLATION COST	121.8	122.1	120.3	128.9	120.3	114.6	121.8
INDIRECT FIELD COST	91.4	91.8	90.2	96.6	90.3	85.9	91.5
FIELD CONSTRUCTION COST	873.0	873.8	878.4	880.9	873.6	806.5	879.8
ENGINEERING SERVICES	54.9	55.1	53.8	56.7	54.3	51.6	54.9
CONTINGENCY	147.1	147.0	150.5	152.5	146.9	135.6	148.4
OXYGEN PLANT	NA	NA	NA	NA	NA	NA	NA
ESCALATION AND INTEREST DURING CONSTRUCTION	123.6	123.7	124.5	125.4	123.6	114.3	124.6
TOTAL CAPITAL COST	1198.6	1199.6	1207.2	1215.5	1198.4	1108.0	1207.7
PLANT OUTPUT (MWe)	1257.4	1248.2	1165.4	1152.8	1272.0	1166.9	1246.6
CAPITAL COST ($/kWe)	953.2	961.1	1035.9	1054.4	942.1	949.5	968.8
COST OF ELECTRICITY: (mills/KwHr)							
CAPITAL COST	30.13	30.38	32.75	33.33	29.78	30.02	30.63
FUEL COST	15.52	15.72	15.86	15.60	15.40	15.37	15.68
OPERATION AND MAINTENANCE	6.40	6.45	6.44	6.66	7.93	6.69	6.44
TOTAL COE	52.05	52.55	55.05	55.59	53.11	52.08	52.75
OVERALL PLANT EFFICIENCY (%)	43.45	42.91	42.53	43.22	43.80	43.89	43.02

*See Table 1.5-2 for details of case variations

Table 4.4-3. Base Case 2 Cost Summary Through COE
(Mid-1978 Dollars, Fuel and O&M Levelized) (Continued)

CASE NO.	2.4	2.4a	2.5	2.6	2.7	2.10	2.11
PRIMARY CHANGE FROM REFERENCE CASE*	NASA Spec.	GE Recalc.	Cs Seed	Super Sonic	8T	1500 MWt	2000 MWt
CAPITAL COSTS: ($ X 10^6)							
MAJOR EQUIPMENT COST	496.4	496.8	483.3	601.5	676.6	290.6	382.9
BOP MATERIAL COST	150.9	152.6	151.2	155.4	153.1	88.7	113.3
INSTALLATION COST	120.8	121.4	120.0	125.2	122.5	78.9	95.5
INDIRECT FIELD COST	90.7	91.1	89.9	94.1	91.9	59.3	71.7
FIELD CONSTRUCTION COST	858.8	861.9	844.4	976.2	1044.1	517.5	663.4
ENGINEERING SERVICES	54.4	54.8	54.2	56.2	55.1	34.0	42.1
CONTINGENCY	144.4	144.7	141.9	167.4	181.9	86.6	111.7
OXYGEN PLANT	NA	NA	NA	NA	NA	NA	NA
ESCALATION AND INTEREST DURING CONSTRUCTION	121.6	122.1	119.7	138.0	147.3	73.4	94.0
TOTAL CAPITAL COST	1179.2	1183.5	1160.2	1337.8	1428.4	771.5	911.2
PLANT OUTPUT (MWe)	1242.3	1255.7	1280.8	1255.6	1293.2	654.5	887.2
CAPITAL COST ($/kWe)	949.2	942.5	905.8	1065.5	1104.5	1087.1	1027.1
COST OF ELECTRICITY: Mills/KwHr)							
CAPITAL COST	30.01	29.79	28.64	33.68	34.92	34.37	32.47
FUEL COST	15.59	15.60	15.13	15.75	15.09	16.14	15.79
OPERATION & MAINTENANCE	6.11	6.07	6.34	6.63	6.30	7.37	6.89
TOTAL COE	51.71	51.46	50.11	56.06	56.31	57.88	55.16
OVERALL PLANT EFFICIENCY (%)	43.26	43.23	44.57	42.81	44.69	41.78	42.70

*See Table 1.5-2 for details of case variations

Table 4.4-3. Base Case 2 Cost Summary Through COE
(Mid-1978 Dollars, Fuel and O&M Levelized) (Continued)

CASE NO.	2.11a	2.12	2.15	2.16	2.16a
PRIMARY CHANGE FROM REFERENCE CASE*	Slag Rej.	Atm Air Htr Comb.	Dual Flow Train	Hot Bottom HTAH, 4 kV/m	Hot Bottom HTAH, 6-5 T
CAPITAL COSTS: ($ X 10^6)					
MAJOR EQUIPMENT COST	382.4	552.0	671.7	431.2	430.6
BOP MATERIAL COST	113.7	160.0	195.3	142.4	142.6
INSTALLATION COST	95.7	133.4	153.3	113.1	113.4
INDIRECT FIELD COST	72.0	100.0	115.1	84.9	84.9
FIELD CONSTRUCTION COST	663.8	945.4	1135.4	771.6	771.5
ENGINEERING SERVICES	42.2	59.0	69.6	51.1	51.1
CONTINGENCY	111.6	162.8	198.6	129.0	128.7
OXYGEN PLANT	NA	NA	NA	NA	NA
ESCALATION AND INTEREST DURING CONSTRUCTION	94.0	134.2	161.4	109.5	109.4
TOTAL CAPITAL COST	911.6	1301.4	1565.0	1061.2	1060.7
PLANT OUTPUT (MWe)	880.7	1221.4	1257.4	1202.9	1178.4
CAPITAL COST ($/kWe)	1035.1	1065.5	1244.6	882.2	900.1
COST OF ELECTRICITY: (Mills/KwHr)					
CAPITAL COST**	32.72	33.68	39.35	27.89	28.45
FUEL COST	16.01	15.76	15.52	14.90	15.21
OPERATION & MAINTENANCE	6.95	6.51	6.40	6.55	6.65
TOTAL COE	55.68	55.95	61.27	49.34	50.31
OVERALL PLANT EFFICIENCY (%)	42.13	42.79	43.45	45.25	44.33

*See Table 1.5-2 for details of case variations

**Based on 65% availability for all cases. Dual flow train (case 2.15) would require 85% availability to match capital cost of case 2.0.

Table 4.4-3. Base Case 2 Cost Summary Through COE
(Mid-1978 Dollars, Fuel and O&M Levelized) (Continued)

CASE NO.	2.16b	2.17	2.18
PRIMARY CHANGE FROM REFERENCE CASE*	Cold Bottom HTAH, 4 kV/m	CAPFB Air Htr Comb.	1300 F Air and Flue Gas to Preheat Combustor
CAPITAL COST: ($ X 10^6)			
MAJOR EQUIPMENT COST	506.1	492.1	491.4
BOP MATERIAL COST	151.9	155.8	144.6
INSTALLATION COST	121.2	123.4	116.8
INDIRECT FIELD COST	90.8	92.6	87.7
FIELD CONSTRUCTION COST	870.0	863.9	840.5
ENGINEERING SERVICES	54.6	55.8	52.4
CONTINGENCY	147.0	145.1	142.6
OXYGEN PLANT	NA	NA	NA
ESCALATION AND INTEREST DURING CONSTRUCTION	123.2	122.5	119.1
TOTAL CAPITAL COST	1194.8	1187.3	1154.6
PLANT OUTPUT (MWe)	1273.4	1281.8	1180.6
CAPITAL COST ($/kWe)	938.3	926.3	978.0
COST OF ELECTRICITY: (mills/KwHr)			
CAPITAL COST*	29.66	29.28	30.92
FUEL COST	15.21	15.93	15.32
OPERATION & MAINTENANCE	6.34	6.33	6.64
TOTAL COE	51.21	51.54	52.88
OVERALL PLANT EFFICIENCY (%)	44.34	42.33	44.03

*See Table 1.5-2 for details of case variations.

Table 4.4-4. Base Case 3 Cost Summary Through COE
(Mid-1978 Dollars, Fuel and O&M Levelized)

CASE NO.	3.0	3.1	3.2	3.4	3.5
PRIMARY CHANGE FROM REFERENCE CASE*	Ref.	I6	2-Stage Cyclone	1100 F Air	8T
CAPITAL COSTS: ($ x 10^6)					
MAJOR EQUIPMENT COST	326.3	350.6	333.6	331.2	494.2
BOP MATERIAL COST	140.3	138.2	139.8	142.5	141.5
INSTALLATION COST	123.1	120.3	122.1	124.2	134.9
INDIRECT FIELD COST	92.3	90.4	91.5	93.1	101.2
FIELD CONSTRUCTION COST	682.0	699.5	687.0	691.0	871.8
ENGINEERING SERVICES	53.4	52.4	53.0	53.9	56.6
CONTINGENCY	106.9	110.8	108.1	108.1	146.4
OXYGEN PLANT	85.6	85.4	85.6	87.0	85.6
ESCALATION AND INTEREST DURING CONSTRUCTION	106.7	109.0	107.4	108.1	133.5
TOTAL CAPITAL COST	1034.6	1057.1	1041.1	1048.1	1293.9
PLANT OUTPUT (MWe)	1089.3	1110.8	1088.8	1098.8	1118.7
CAPITAL COST ($/kWe)	949.8	951.7	956.2	953.9	1156.6
COST OF ELECTRICITY: (mills/Kw/Hr)					
CAPITAL COST	30.03	30.09	30.23	30.15	36.56
FUEL COST	15.72	15.68	15.72	15.85	15.30
OPERATION & MAINTENANCE	7.13	9.52	7.12	7.13	7.00
TOTAL COE	52.88	55.29	53.07	53.13	58.86
OVERALL PLANT EFFICIENCY (%)	42.91	43.02	42.89	42.55	44.07

*See Table 1.5-2 for details of case variations

subsystem results in a small saving in COE. Net gain in plant output with the 7T field (Case 1.3) was insufficient to counterbalance increased magnet cost and differences in combustor design (Cases 1.4 and 1.4a) had little effect on cost.

Variations in MHD generator and system configuration were concentrated in Base Case 2 (pressurized HTAH) and one of these, Case 2.16, gave the best COE of all the cases examined, as well as an increase in efficiency relative to Reference Case 2.0. Several cases from Table 4.4-3 illustrate singly and together the improvements which led to the Case 2.16 results. Case 2.16a shows that part of the improvement over Case 2.0 which resulted from additional thermodynamic regeneration made possible by a higher temperature at the bottom of the HTAH array and the concomitant reduction in the height and therefore the cost of the ceramic heat exchangers.

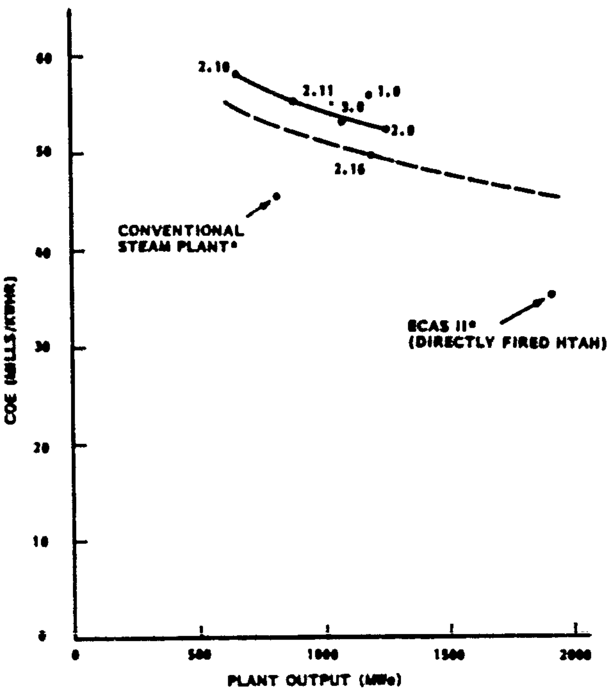
Case 2.16 also utilized improved MHD generator performance obtained by specifying a uniform transverse electric field of 4 KV/m (Case 2.16b) and computing the required magnetic field distribution. A peak magnetic field greater than 6T results, but the field distribution is such that structural containment requirements are not much more severe than the nominal 6T tapering to 5T case. Also, while not included in this phase of the study, indications are that essentially the same MHD generator performance can be reached by a combined constraint of 6T peak magnetic field and 4 KV/m transverse electric field. The magnet for Case 2.16 was therefore assumed to cost the same as the 6T magnet for Case 2.0.

By itself, the hot bottom approach, Case 2.16a, results in a 1.7 mill/KWhr saving over the reference case, while the 8T magnet, costing 2.5 times the 6T magnet (Case 2.7), adds 4.3 mills/KWhr to COE. The margin of improvement from Case 2.16a to Case 2.16 is therefore dependent on a more precise evaluation of incremental magnet cost, if any.

Some coal and combustion system variations, all involving moderate slag carryover, were also examined. The best pressurized moving bed gasifier case was that which assumed split stream slagging pressurized moving bed (S^3PMB), supplemented with direct coal firing. While the differences are not large, it appears that the reference case two-stage cyclone approach is superior. Use of Illinois #6 coal (Case 2.1) improves efficiency and reduces cost because of the reduced moisture and additional heating value but this improvement is partly offset by the penalty for additional sulfur removal*.

The relationship between the cost of electricity and power plant size was explored in Cases 2.10, 2.11 and 2.0. As might be expected, COE varies inversely with plant size due to a combination of increased unit capital cost and reduced efficiency. COE for these three cases is plotted in Figure 4.4-1, along with the value for Case 2.16. A curve parallel to the reference case size variation curve has been plotted through the point corresponding to Case 2.16

* Here dry scrubbing applies only to the HTAH combustion gas stream and the COE increments over the tabulated values are 0.5 mills/KWhr capital and 1.4 mills/KWhr levelized operating cost for Case 2.0s. The comparable figures for Case 2.1 are 0.8 mills/KWhr and 1.8 mills/KWhr, respectively.



* COE from the original studies has been recalculated using PSPEC rules for contingency and for escalation and interest during construction. Costs have been escalated to mid-1978 dollars. Fuel and O&M costs have been levelized.

Figure 4.4-1. Cost of Electricity as a Function of Plant Size and Type

and extrapolated to cover the range of power plant sizes being considered for commercial operation. The costs of electricity for the other two reference cases, as well as the open cycle MHD plant and the conventional steam plant from ECAS II[9] are shown for comparison. The OCMHD plant was 4000 MWt and assumed the use of direct fired air heaters. The conventional steam plant analysis was on the assumptions of wet lime stack gas scrubbers and stack gas reheated to 175 F[10].

COE is calculated from data in original studies escalated to mid-1978 dollars at 8% and using PSPEC rules for contingency and for interest and escalation during construction. Fuel and operation and maintenance costs were levelized using PSPEC ground rules. Costs are higher than, but close to, those of the conventional plant. However, the assumed fuel cost is relatively low and was assumed to increase only at the general inflation rate during the life of the plant. In addition, the conventional plant cannot meet the emission standards to which the MHD plants were designed and emission standards, which, along with availability, may well be the most important criteria to be met.

In making the estimate for the dual power train (Case 2.15) the HTAH cost was multiplied by 1.05 to account for the additional piping and valves required to supply the second combustor.

Also the magnet cost was multiplied by 1.95 on the assumption that total duplication of the magnet support system would not be needed. As indicated in Table 4.4-3, the increase in COE for the dual power train approach is more than 9 mills/KWhr. However, the effect of presumably greater availability was not included. In order to match the capital cost of the Reference Case (Case 2.0), availability would have to be 85%, as opposed to the 65% assumed for all cases.

As in Base Case 1, parametric variations for Base Case 3, Table 4.4-4, made some minor differences in cost and performance but no substantial changes from the reference case were obtained except for the 8T magnet, Case 3.5, where a 1-point efficiency improvement was insufficient to overcome the larger magnet cost. Enrichment to 40 lb O_2 per 100 lb air (equivalent to about 42 mole % oxygen in the mixture) was chosen on the basis of selecting the highest net MHD generator output (gross MHD less power for main compressors and O_2 production) from among calculations for 20, 30 and 40 lb/100 lb air. Overall cost optimization during conceptual design is expected to result in a lower degree of O_2 enrichment.

SECTION 5

DEVELOPMENT ISSUES

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DEVELOPMENT ISSUES

Since completion of the ECAS and the ETF conceptual design studies, there has been widespread recognition that a consideration of system implications and component interactions is essential to the orderly development of OCMHD. The combustor and MHD generator taken together remain the central issue in achieving the performance, reliability and desirability essential to commercialization of the MHD power generation. Other subsystems also require development at more than the single component level, in particular the Heat Recovery/Seed Recovery and High Temperature Air Heater Subsystems.

Dynamic analysis, considering start-up, shut-down and load change as well as oscillations under nominally steady state conditions, while not within the scope of this study, is an area which must also be considered during flow train development. A capability for predicting such interactions could be important in the design of scaled up components and the development of control strategies and operating procedures.

5.1 COMBUSTOR

This study was limited to systems with zero to moderate slag carryover, which puts an additional constraint on combustor performance beyond those of reaching conductivity, uniformity, temperature, reliability and durability goals.

5.1.1 MODERATE SLAG CARRYOVER COMBUSTORS

The moderate slag carryover systems all depend on some form of cyclonic separation and must either perform this separation at temperatures low enough to inhibit slag vaporization or depend on kinetic effects to maximize separation before vaporization has time to occur. To the level of analysis undertaken herein, the expected performances of the single-stage vortex and the two stage slagging cyclone are essentially indistinguishable and the issues in further development are:

1. Achievement of performance goals:

   Can specified conductivity, temperature and slag rejection be reached?

2. Uniformity

   Can a plasma with sufficiently uniform properties (temperature, conductivity) be delivered to the MHD generator?

3. Scaling:

How do slagging characteristics and heat loss change with size?

What size modules are optimum and how can the modules be manifolded to minimize heat loss?

4. Stability:

Stead-state performance analysis indicates that subsonic operation of the MHD generator is clearly superior from a performance standpoint, but the combustor and generator are then gas dynamically coupled. Oscillations resulting from combustor-generator interactions are likely, and are a particularly critical scale-up consideration.

5. Durability:

Will the combustor operate reliably for a lifetime satisfactory for electric utility operation?

5.1.2 ZERO SLAG CARRYOVER COMBUSTORS

The advantages and drawbacks of zero slag carryover operation were addressed in the GE ETF study[1]. The advantages include electrical isolation between combustor stages (so that coal and oxidizer feed and slag removal systems can operate at ground potential) plus simplification of seed recovery by elimination of the need for slag/seed separation. However, hot ceramic electrodes require further development and a suitable combustor must be identified. Some form of gasifier/combustor is required, and this has the potential further advantage of drawing on technology already under development.

An adaptation of the Foster Wheeler fluidized bed concept was considered during the ETF study, but the dual constraints of bed temperature low enough for solid ash removal and gas heating value sufficient to reach final plasma temperature resulted in very severe first stage fuel/oxidizer ratio and inert bed circulation requirements.

A slagging gasifier relaxes both of the above constraints and results in potential operation at conditions comparable to existing gasifiers. As described in Section 3.1.2, a variant of the slagging moving bed gasifier concept was studied for PSPEC. The concept utilizes two gas streams, one CO rich stream is used for the MHD combustors and the other, containing most of the hydrocarbons and moisture, is used for the indirectly fired HTAH subsystem. Existing test data with this type gasifier is for oxygen/steam blow operation at relatively high pressures ($p \sim 25$ atm). Operating data for air blown only operation at pressures, in the range of 5 to 10 atmospheres, would be required to verify projected MHD operating conditions and to determine the maximum preheat air temperature capability of this gasifier operating air blown for oxygen enrichment without steam might also be considered. Split stream operation requires some hardware development and will probably require integration with a separate application such as a petrochemical plant.

5.2 MHD GENERATOR

Substantial progress has been made in achieving long life in electrode and channel operation under realistic local conditions of temperature, chemistry, current density and voltage gradient. Still, achievement of the plant efficiency range shown in this study is heavily dependent on reaching the output power levels predicted by the quasi-one dimensional analysis of the combustors and the MHD generator.

Scaling of predictions over a range from presently contemplated 20 to 50 MWt testing through the ETF-size range of 500 MWt to full scale plants in the 1200 to 2800 MWt range requires an assessment of the relative importance of many effects not fully accounted for with the performance model used in this study. Thus, include electrode voltage drop with the influence of electrode surface temperature and slag deposition and end effects.

Verification of this performance requires testing at a scale and magnetic interaction sufficient that losses are not dominant and enthalpy extraction becomes a significant fraction of thermal input power. Constant electrical stress design can, in principle, significantly increase generator output but testing will be required to prove the concept, with regard to both power output and electromechanical design.

Since magnet cost is an important contribution to overall plant cost, channel construction must emphasize minimization of the ratio of warm bore to gas dynamic cross-section. Ability to withstand the axial voltage gradient may require use of non-electrical conducting structural materials probably of the reinforced epoxy or polyimide class. This, in turn, requires a coolant system capable of maintaining temperatures below 500°F. Isolation of generator coolant from the primary boiler feedwater loop will probably also be required, both to maintain feedwater purity and to simplify channel replacement.

Power takeoff and inversion also require development effort. The conclusion was reached in both the GE[1] and AVCO[2] ETF studies that insufficient data were available to make an effective choice between the Faraday and diagonal concepts*. As in the case of one-stage versus two-stage combustors, the final selection will depend on factors other than predicted design performance. These include:

1. Fabrication methods
2. Construction and maintenance cost
3. Cooling requirements
4. Capability and need for control of individual electrode pairs
5. Off-design performance
6. Power takeoff and inversion cost

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* Linear channels only were considered. The disc concept is the subject of a separate study now in progress[3].

5.3 MAGNET

There appears to be less uncertainty regarding the feasibility of extrapolation of current superconducting magnet technology to baseload size than for some other parts of the MHD plant. At maximum field of 6 to 7 Tesla, niobium-titanium superconductor is adequate and this study has shown that there is little to be gained by attempting to operate at higher field strengths.

However, while feasibility is not in question, there is substantial uncertainty regarding the cost of this most expensive single component of the MHD/steam power plant. In addition to warm bore specification, cost is dependent on design, fabrication methods and shipping limitations because the cost of the structure is an order of magnitude greater than the cost of the conductor.

5.4 HEAT RECOVERY/SEED RECOVERY SYSTEM

Identification of Kraft recovery boiler technology in the AVCO/Combustion Engineering ETF study[2] was a significant forward step in design of the steam generator component for a moderate slag carryover MHD/steam power plant. The presently planned 20 MWt[4] experimental study should be directed toward obtaining the design data necessary to construct the HRSR at ETF and baseload scales. The primary issues, all concerned with the gas side, are:

1. NO$_x$: Can NO$_x$ emission standards be met by controlling the time-temperature profile at plasma exiting the MHD diffuser. This problem is particularly acute at the 20 MWt size because control of cooling rate is more difficult at the 20 MWt size than at larger sizes and a ceramic lining of the furnace, which is unnecessary at 300 MWt and larger sizes will be required. In spite of the extra difficulty, it will probably be necessary to actually demonstrate effective NO$_x$ reduction at the 20 MWt size.

2. Seed/Slag Separation: The HRSR design must be capable of separating slag from seed sufficient for economic overall plant operation. Costing for this study was done on the assumption of 95% seed recovery but some additional parametric analysis to determine sensitivity to a range of recovery levels is appropriate.

3. Gas Emissivity: A major uncertainty in HRSR design is the influence of particulates on the radiant emission characteristics of the plasma in the Radiant Furnace. The particulate emission factor F_E was set equal to 1.1 (for no particulate, $F_E = 1.0$) for PSPEC analysis but experimental data are required.

4. Afterburn: Initial combustion with 85 to 90% stoichiometric oxidizer produces both maximum plasma temperature and minimum NO$_x$. Combustion is therefore completed in the HRSR subsystem. Analysis indicates that this can be accomplished in a temperature range low enough to avoid additional NO$_x$ formation but high enough for complete conversion to CO but development testing is necessary.

A final issue in the HRSR system is one of materials survivability. The 2000 hour test life contemplated for the 20 MWt HRSR experiment is insufficient to establish long term survivability. Detailed economic trace-offs with reliable materials data will be required to optimize the HRSR system.

5.5 SEED REPROCESSING

Both alternatives for seed reprocessing which were examined in this study are technically feasible, but the formate process is clearly more economical than the electrochemical, in both capital and operating cost. The Hooker Chemical Company recommendation of using a slurry rather than a saturated solution in the formate process has a distinct system advantage in reducing thermal energy needed for water evaporation.

Reactor size was established on the basis of extrapolation from a similar sodium sulfate process (3 times the $NaSO_4$ reactor size was assumed) but experimental evaluation is needed. A relatively modest, separate, oxygen blown, coke gasifier is required. The energy requirements are such that the plant is very nearly self-contained and a conceptual design would probably be on that basis. Given a reasonable storage capacity for both spent and reprocessed seed, if the seed reprocessing plant is self-contained, it can operate at the average capacity of the power plant independent of load fluctuations.

Economics appear to favor "once-through" processing to calcium sulfate rather than recovery of elemental sulfur. The end product is stable for disposal provided the spent seed is fully converted from sulfite to sulfate prior to reconversion to formate.

5.6 AIR HEATERS

Of the cases considered in this study, those with indirectly fired high temperature air heaters with pressurized reheat supplying 3000°F preheated air are the most efficient and cost effective. The major development issues for such a system are:

1. Efficient Combustion System: The regeneratively air cooled high slag rejection cyclone combustor now under development at GE, combined with dry scrubbing (possibly with coal beneficiation) is recommended. This combustor has performed well at 1 atmosphere and tests at up to 4 atmospheres are already planned.

2. Slag Carryover: Complete elimination of slag/ash carryover from a cyclone combustor does not appear feasible. Provision must be made, therefore, for periodic removal of slag and ash which may collect in the bed. Recent data from Montana State University[5] and from FluiDyne Engineering Co.[6] suggest that it will be feasible to melt out slag by periodically heating the bottom of the air heater to a temperature above slag melting.

3. Gas Turbine Drive for Pressurization: Gas turbines have proved sensitive to particulate matter, especially alkali metals salts when used with coal combustion products. However, in the air heater application inlet temperature is approximately 900°F and data from locomotive gas turbine development[7] and other applications indicates that this temperature is low enough to assure sufficient turbine life.

The inherent thermodynamic advantage of directly fired air heaters has been partially overcome by careful integration of the recuperative air heaters. There are also several practical advantages of the indirectly fired system in addition to the obvious one of avoiding the necessity

to survive the slag/seed laden plasma from the MHD power train. These advantages lie in the pressurization of the combustion process in the indirectly fired case. With pressurized reheat the high temperature valves need not withstand more than a nominal pressure differential and there is no need for pressurization or pressure letdown stages in the air heater cycle.

5.7 O_2 PLANT

An obvious advantage of O_2 enrichment without a regenerative air preheat system is that development of the latter can be delayed at least until after the MHD flow train has demonstrated its value.

Since the O_2 plant is essentially an off-the-shelf package the only development issues relate to selection of compressor and other components for maximum cost effectiveness and to possibilities (or desirabilities) of integration with the rest of the system.

SECTION 6

SUMMARY AND CONCLUSIONS

SECTION 6

SUMMARY AND CONCLUSIONS

6.1 GENERAL

Figure 6.1-1, repeated from Section 2, summarizes the results of performance analyses completed in this phase of the PSPEC study. As indicated in Figure 6.1-1, calculated coal pile to bus bar efficiencies fall in the range of 42 to 46%. This range is consistent with "black box" type estimates of efficiencies relative to those for the directly fired system considered in ECAS II. A reduction in performance associated with more realistic estimates of some losses, such as those in the combustor, was compensated for by improvements in plant integration, including partial regeneration, better steam plant performance and reduced energy for seed reprocessing. The combustor and generator must be considered together because, since exit pressure is fixed, generator performance determines operating pressure and both components are sensitive to pressure.

Differences in both performance and cost determination in this study, for individual parametric variation, while large in absolute Megawatts and dollars are quite small on a "per unit" basis. Nevertheless, the parametric cases showed sufficient variation to suggest combinations better suited for early commercial design. Calculations have been carried to a precision sufficient to avoid round-off error for case by case comparison and are not meant to imply that absolute accuracy.

To the precision of the calculations and the assumptions on which they are based, there is no significant difference in performance between a single-stage combustor with 70% slag rejection and a two-stage combustor with 85% slag rejection. This is the result of several compensating phenomena including the effect of operating pressure on heat loss and the influence of slag on plasma conductivity. ETF size modules were assumed, neglecting manifolding in the single stage case but accounting for manifold losses to a single second stage combustor in the two-stage case. Slagging gasifiers can produce a high-temperature, high-conductivity slag-free plasma but complete and effective use of the moisture and volatiles laden "top gas" stream from the S^3PMB (See Section 3.1.2) requires a separate application such as petrochemical production.

Analysis also indicates that MHD generator performance can be improved by tailoring the magnetic field so as to produce a constant electrical stress in the channel. Estimating the cost of construction of baseload size magnets requires substantial extrapolation from existing information, particularly if an 8 - 7 Tesla field is specified. The stronger field offers only a modest ($\sim$ 1.2 percentage points) theoretical improvement and is not warranted for any early commercial plant. Structural containment is the dominant cost element and a field between 6 and 7 Tesla near the upstream end where bore size is smaller tailored to approximate the constant electrical stress condition is recommended.

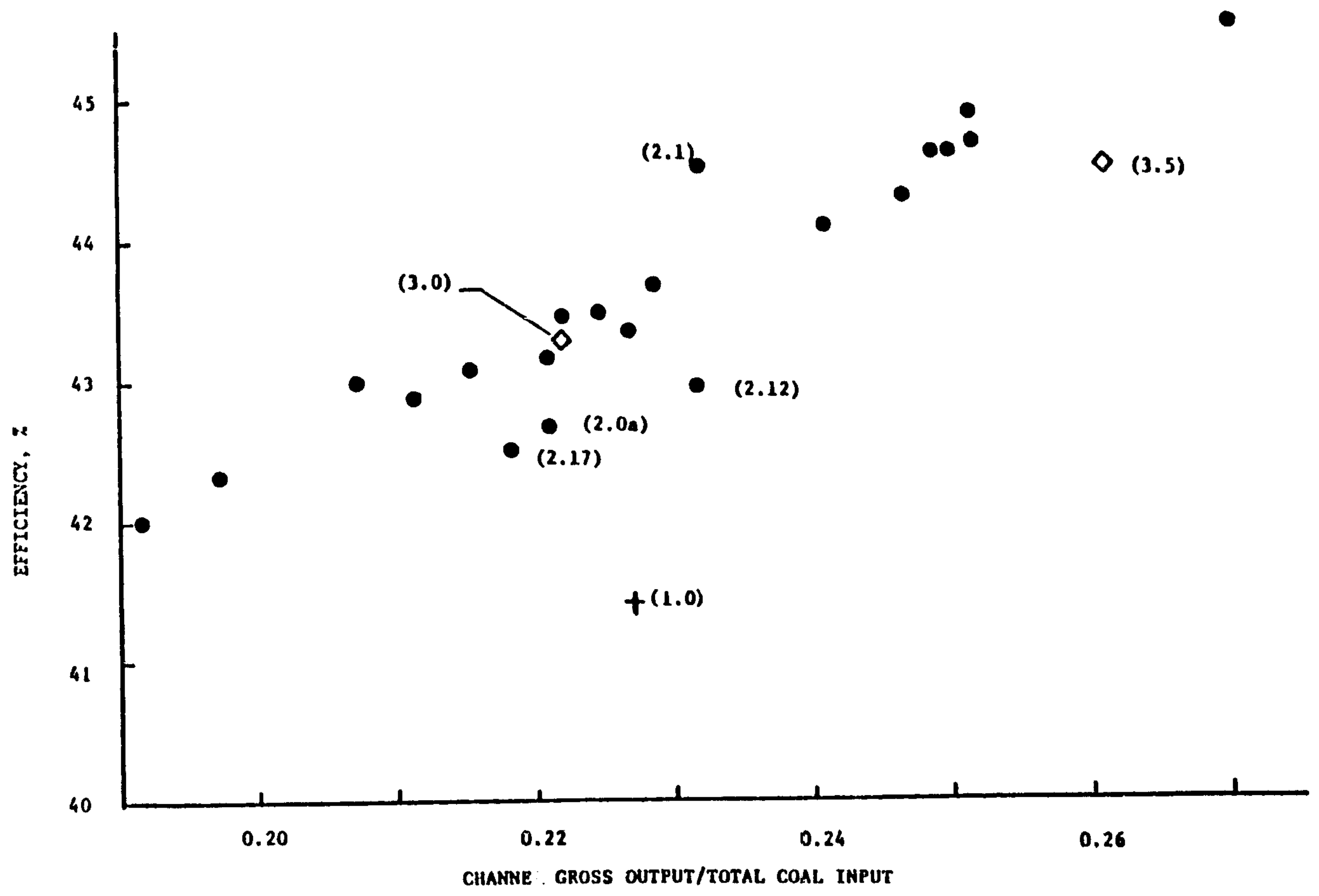


Figure 6. 1-1. Relationship of Plant Efficiency to Channel Output

6.2 SUMMARY OF BASE CASE RESULTS

6.2.1 BASE CASE 1

This case considered an indirectly fired HTAH delivering 2700 F air with an atmospheric pressure gasifier for HTAH reheat. The efficiency range was 41.4% to 42.9% including flue gas desulfurization with the dry scrubbing process for both the HTAH and MHD combustion products. The HTAH penalty is significant in this system and there is no net gain from O_2 enrichment. These systems had the highest cost of electricity and are not recommended for development.

6.2.2 BASE CASE 2

This case considered an indirectly fired HTAH delivering 3000 F air. Pressurized reheat was assumed for all but one parametric variation. Plant efficiency ranged from 42.0% to 45.6% including seed reprocessing by the formate process. Both gasifiers and cyclone (or vortex) combustors were examined. The best efficiency, Case 2.16, was obtained with a hot bottom (1300 F input air) HTAH and an MHD generator operating in a magnetic field tailored to keep electrical stress constant. Case 2.16 had the lowest cost of electricity and is recommended for conceptual design.

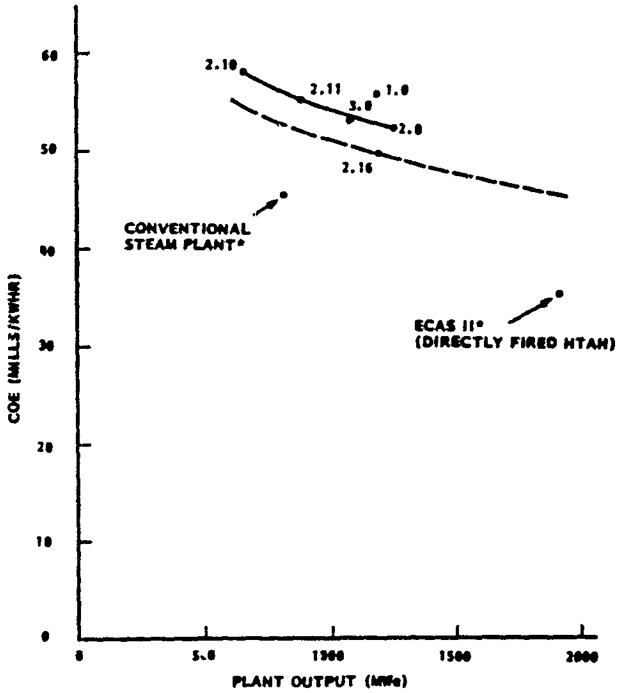
6.2.3 BASE CASE 3

This case considered recuperative air preheat in the range of 1100 F to 1300 F, combined with O_2 enrichment to 42% by volume (See Section 1.3). In the absence of an HTAH with its high thermal losses, O_2 enrichment can yield efficiencies comparable to Base Case 2. Cost of electricity is higher than for Base Case 2 but improvement is possible with further tradeoffs of cost versus enrichment level and there may also be some opportunity to integrate the O_2 plant with the MHD/steam plant. Development of the HTAH can be postponed and the O_2 enrichment concept is also recommended for conceptual design.

6.3 COSTING

Figure 6.3-1 repeated from Section 4.0 summarizes costing results. The increase in capital cost relative to the ECAS II directly fired HTAH plant shown in Figure 6.3-1 is primarily a result of higher estimates for the magnet and HTAH (or O_2 plant) subsystems. There are other capital cost differences, most notably the low temperature recuperative air heater cost is substantially lower. Recuperative air heater cost reduction is the result of a change in concept to hanging panels with headers which are not exposed to the high temperature gas plus utilization of high velocity on the air side to keep metal temperature close to the air temperature. Changes in absolute levels of costs are partly due to a change in base year from 1975 to 1978 and partly because costs have been levelized to present a more accurate comparison with other, less efficient systems.

Note added in proof: The O_2 enrichment concept has been selected for conceptual design



* COE from the original studies has been recalculated using PSPEC rules for contingency and for escalation and interest during construction. Costs have been escalated to mid-1978 dollars. Fuel and O&M costs have been levelized.

Figure 6.3-1. Cost of Electricity as a Function of Plant Size and Type

The reference cases for the three base cases (Cases 1.0, 2.0 and 3.0) have slightly different outputs because, for analytical convenience, input thermal power to the MHD power train was held constant at 2800 MWt. Plant size variation was considered in Base Case 2 (1500 MWt and 2000 MWt to MHD power train) and a size cost trend was established which indicates potential COE gains to 2000 MWe. The dashed line through Case 2.16 was drawn parallel to the size trend line.

For comparison, both the conventional steam plant studied as part of ECAS[1] and the ECAS II[2] directly fired HTAH results are also plotted. Costs have been recalculated consistent with the 1978 base year and the levelizing algorithm used in this study.

Costs are higher than, but close to, those of the conventional plant. However, the assumed fuel cost is relatively low and was assumed to increase only at the general inflation rate during the life of the plant. In addition, the conventional plant cannot meet the emission standards to which the MHD plants were designed and, emission standards, along with availability, may well be the most important criteria to be met.

SECTION 7

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SECTION 7

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APPENDIX A

DICTIONARY OF NODE NAMES

APPENDIX A

DICTIONARY OF NODE NAMES

Nodes shown on system diagrams in this report are defined below. They were given alphanumeric designations intended for use in system code OCSYS as well as on the diagrams.

```
AH1-----AIR HEATER #1  HEATS OUTPUT OF THE MAIN COMPRESSOR TO  650K
AH2-----AIR HEATER#2 HEATS THE MAIN COMBUSTION AIR TO FINAL TEMPERATURE
AM-----AMBIENT STATE POINT NODE
AMA-----PREHEAT COMBUSTOR AIR STREAM
BFP-----BOILER FEED PUMP
BFPT-----BOILER FEED PUMP TURBINE
CAPFB-----CHEMICALLY ACTIVE PRESSURIZED FLUIDIXED BED GASIFIER
CB-----COMBUSTOR
CB1-----COMBUSTOR FIRST STAGE
CB2-----COMBUSTOR SECOND STAGE
CD-----TURBINE CONDENSER
CDR-----COAL DRYER
CDR2-----COAL DRYER FOR PREHEAT COMBUSTOR
COLM-----MAIN COAL STREAM
COLP-----PREHEAT LOOP COAL INPUT STREAM
CP-----CONDENSATE PUMP
CPI-----TOTAL COAL STREAM AS USED IN COAL DRYER
DA-----DEAERATOR
DF-----DIFFUSER
ECH-----HIGH TEMPERATURE/PRESSURE ECONOMIZER
ECH2-----HIGH TEMPERATURE/PRESSURE ECONOMIZER #2
ECL-----LOW TEMPERATURE/PRESSURE ECONOMIZER
ESP1-----ELECTRO STATIC PRECIPITATOR #1
ESP2-----ELECTRO STATIC PRECIPITATOR #2
ESP3-----ELECTROSTATIC PRECIPITATOR #3
FGD-----FLUE GAS DESULFURIZATION
FOF1-----FINAL OXIDATION FURNACE----------SECTION 1
FOF2-----FINAL OXIDATION FURNACE----------SECTION 2
FW1-----FEED WATER HEATER #1 (HIGHEST TEMPERATURE)
FW2-----FEED WATER HEATER #2
FW22-----DRAIN OF FEED WATER HEATER #2
FW4-----FEED WATER HEATER #4
FW42-----DRAIN OF FEED WATER HEATER #4
NOTE-----FW3 IS NOT USED
FW5-----FEED WATER HEATER #5
FW52-----DRAIN OF FEED WATER HEATER #5
FW6-----FEED WATER HEATER #6
FW62-----DRAIN OF FEED WATER HEATER #6
GEN-----MAIN GENERATOR
GMA-----GAS MIXER "A"
GMB-----GAS MIXER "B"----------NODE FOR JOINING 2 GAS STREAMS
GMC-----GAS MIXER "C"
GMD-----GAS MIXER "D"
GSA-----GAS SPLITTER "A"
GSA2-----GAS SIDE STREAM CREATED BY GAS SPLITTER "A"
GSB-----GAS SPLITTER "B" NODE FOR DIVIDING A GAS STREAM
GSB2-----THE "SIDE" STREAM CREATED WITH GSB
GSC-----MAIN STREAM OF A GAS SPLITTER "C"
GSC2----- "SIDE" STREAM CREATED BY GAS SPLITTER "C"
HPT-----HIGH PRESSURE TURBINE
HPT2----- HIGH PRESSURE TURBINE SHAFT LEAKAGE NODE
HPT3-----HIGH PRESSURE TURBINE SHAFT LEAKAGE NODES
HPT4-----HIGH PRESSURE TURBINE SHAFT LEAKAGE NODE
HPT5-----HIGH PRESSURE TURBINE SHAFT LEAKAGE NODE
HPT6-----OUTPUT OF CONTROL STAGE AND LEAKAGE TO SHELL
HTAH-----HIGH PRESSURE AIR HEATER
IDF-----INDUCED DRAFT FAN
INV ----INVERTER
IPP-----INTERMEDIATE PRESSURE PUMP
```

```
LPAH-----LOW PRESSURE AIR HEATER
LPT-----LOW PRESSURE TURBINE
LPT1  (UNUSED NODE)
LPT2-----LOW PRESSURE TURBINE EXTRACTION PORT
LPT3-----LOW PRESSURE TURBINE EXTRACTION PORT
LPT4-----LOW PRESSURE TURBINE EXTRACTION PORT
LPT5-----LOW PRESSURE TURBINE EXTRACTION PORT
LPTX-----LOW PRESSURE TURBINE LAST STAGE EXHAUST
MC-----MAIN COMPRESSOR
MD-----MHD CHANNEL
NZ-----NOZZLE
OX-----NODE AT WHICH FINAL OXIDATION AIR IS INJECTED
PGRB-----PREHEAT GAS RECIRCULATION BLOWER
PHB-----PREHEAT BLOWER
PHC-----PREHEAT COMPRESSOR
PHCB-----PREHEATER COMBUSTOR
PHT-----PREHEAT TURBINE
RB1-----FIRST SECTION OF RADIANT BOILER
RB2-----SECOND SECTION OF RADIANT BOILER
RHH-----HIGH TEMPERATURE SECTION OF REHEATER
RHL-----LOW TEMPERATURE SECTION OF REHEATER
RHT-----REHEAT TURBINE
RHT2-----RHEAT TURBINE SHAFT LEAKAGE NODE
RHT3-----REHEAT TURBINE STEAM EXTRACTION PORT
RHT4-----REHEAT TURBINE SHAFT END LEAKAGE NODE
SAH-----SECONDARY AIR HEATER
SCAB-----SECONDARY COMBUSTION AIR BLOWER
SHH-----HIGH TEMPERATURE SECTION OF THE SUPERHEATER
SHL----- LOW TEMPERATURE SECTION OF THE SUPERHEATER
SPE-----SHAFT PACKING EXPANDER
SPE2-----SHAFT PACKING EXPANDER PORT #2
SPMB-----SLAGGING PRESSURIZED MOVING BED GASIFIER
S3PMB-----SPLIT STREAM SLAGGING PRESSURIZED MOVING BED GASIFIER
SSR-----STEAM SEAL REGULATOR
SSR2-----STEAM SEAL REGULATOR PORT #2
SSR3-----STEAM SEAL REGULATOR PORT #3
STK-----PLANT EXHAUST STACK
STP-----STEAM PLANT
STPE-----ELECTRICAL OUTPUT NODE OF THE STEAM POWER PLANT
SX1-----MAIN COMBUSTOR SLAG REJECTION NODE
TP-----TOTAL PLANT
VAL-----THROTTLE VALVE
VAL1-----THROTTLE VALVE LEAKAGE #1
VAL2-----THROTTLE VALVE LEAKAGE #2
VAL3-----THROTTLE VALVE LEAKAG#3
WMA-----WATER MIXER"A" (JUNCTION OF 2 STREAMS)
WMB-----WATER MIXER "B"
WMC-----WATER MIXER "C"
WMD-----WATER MIXER "D"
WME-----WATER MIXER "E"
WMF-----WATER MIXER "F"
WMG-----WATER MIXER"G"
WSA-----WATER SPLITTER"A" (USED AS AN INACTIVE COUPLING NODE)
WSE-----WATER SPLITTER "E"
WSE2-----SECOND PORT OF WATER SPLITTER "E"
WSE3-----THIRD PORT OF WATER SPLITTER "E"
```

APPENDIX B

SYSTEM DIAGRAMS AND ENERGY FLOW SUMMARIES

APPENDIX B

SYSTEM DIAGRAMS AND ENERGY FLOW SUMMARIES

Appendix B contains tabulations of overall efficiencies, a set of system diagrams with state points and a complete set of energy flow summaries. System diagrams in most cases apply to several cases and state points are tabulated only for those cases for which a system code balance was run. An energy flow summary was prepared for all cases considered in the study.

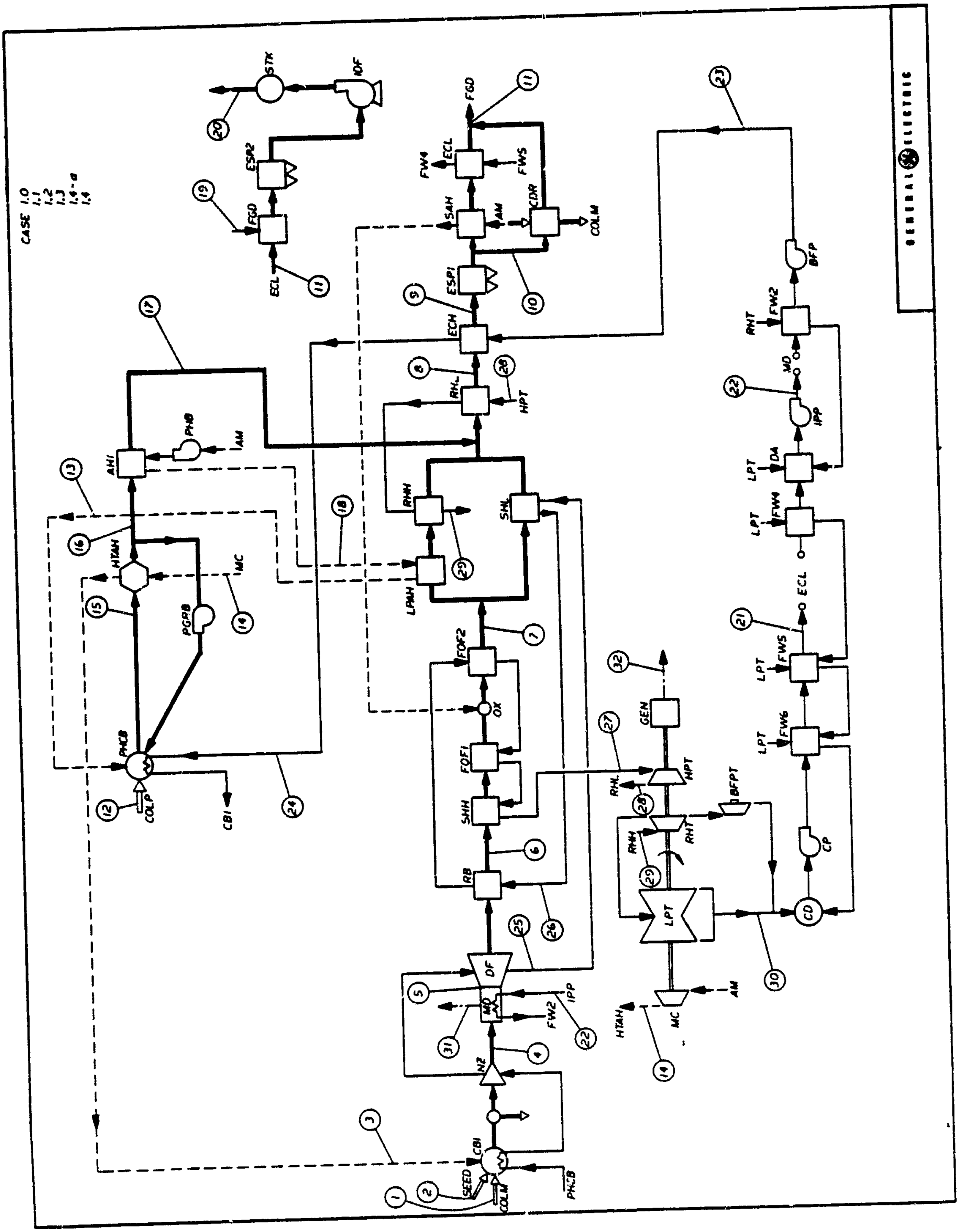
BASE CASE 1

INDEX

Case Number	Overall Efficiency	System Diagram	State Points	Energy Flow Summary
1.0	✓	✓	✓	✓
1.1	✓	✓		✓
1.2	✓	✓		✓
1.3	✓	✓		✓
1.4	✓	✓		✓
1.4A	✓	✓		✓

Overall Plant Efficiency
Base Case 1

Case No.	Parameter Variation	Efficiency, %
1.0	Reference Case	41.41
1.1	Illinois #6 Coal	42.90
1.2	Air Only	41.73
1.3	(7-6) Tesla Magnetic Field	42.21
1.4	Single-Stage Combustor 85% Slag Rejection	41.76
1.4a	Single-Stage Combustor 70% Slag Rejection	41.50


CASE 1.0
1.1
1.2
1.3
1.4-a
1.4
STK
IDF
ESP2
FGD
ECL
FW4
SAH
AM
CDR
COLM
FW5
ESP1
ECH
RHL
HPT
AH1
PHB
HTAH
MC
PGPB
LPAH
RHH
SHL
FOF2
OX
FOF1
SHH
RB
DF
MD
FW2
IPP
NZ
CBI
SEED
PHCB
COLP
GEN
RHT
BFPT
LPT
CD
CP
FW6
DA
BFP
GENERAL ELECTRIC

STATE POINTS, CASE 1.0

Location	T(F)/(K)	P(psia)	M(Kg/S)	E (MW)
1	220/378	-	79.40	2042.0
2	220/378	-	14.28	0.93
3	2700/1756	168.7	458.0	766.3
4	4800/2923	153.4	544.08	2638.8
5	3670/2295	17.60	544.08	1842.7
6	2900/1866	13.94	544.08	1474.5
7	1833/1274	13.82	643.65	874.5
8	762/679	13.66	964.66	599.9
9	513/540	13.56	964.66	449.2
10	513/540	13.56	233.65	109.7
11	336/442	13.43	967.47*	347.3
12	77/298	-	40.33	838.5
13	1100/867	20.0	189.9	116.2
14	688/638	174.9	458.0	165.5
15	3367/2126	17.97	371.66	927.7
16	1100/867	16.17	321.01	242.7
17	774/685	16.12	321.01	174.1
18	728/660	20.0	189.9	73.2
19	59/288	13.43	31.93	0.0
20	221/378	14.75	999.4	363.1
21	190/361	157	544.4	-
22	301/423	400	597.3	-
23	427/493	3900	597.3	-
24	527/548	3880	597.3	-
25	655/619	3670	597.3	-
26	715/653	3610	597.3	-
27	1000/811	3500	597.3	-
28	604/591	768	586.9	-
29	1000/811	691	591.8	-
30	106/314	2.3"Hg	448.2	-
31	-	-	-	654.2
32	-	-	-	644.3

*Includes Coal Drying Moisture

ENERGY FLOW SUMMARY

Case Number: 1.0

Changes from Reference Case: Reference Case

Power In (MW)

Raw Coal to Air Heater Subsystem		838.5	
Raw Coal to Main Combustor		2033.8	
Total Power In			2872.3

Power Out (MW)

Electric Power

MHD Generator		654.2		
Steam Turbines				
Shaft Power	824.7			
Main Compressor	-166.3			
Generator Loss	- 14.1			
Net Steam Power		644.3		
Plant Internal Power				
Inverters	- 9.8			
O_2 Plant	- 32.7			
Electrical Auxiliaries	- 66.5			
Internal Power		-109.0		
Electric Power Out			1189.5	

Thermal Power

Condensate Heat Rejection	1093.0		
Stack Loss	316.6		
Coal Drying	48.1		
Rejected in Solid Waste	11.1		
Internal Electric Power Not Regen.	84.5		
Other Losses to Ambient	130.4		
Thermal Power Out		1683.7	
Total Power Out			2873.2
Unaccounted For			-0.9

Efficiency

Excluding Seed Reprocessing	41.41%
Including Seed Reprocessing	N/A

C-3

ENERGY FLOW SUMMARY

Case Number: 1.1

Changes from Reference Case: Illinois #6 Coal

- -

Power In (MW)

Raw Coal to Air Heater Subsystem			793.6	
Raw Coal to Main Combustor			2036.3	
Total Power In				2829.9

- -

Power Out (MW)

Electric Power				
MHD Generator		691.6		
Steam Turbines				
Shaft Power	812.6			
Main Compressor	-169.8			
Generator Loss	- 13.8			
Net Steam Power		629.0		
Plant Internal Power				
Inverters	- 10.4			
O_2 Plant	- 32.7			
Electrical Auxiliaries	- 63.4			
Internal Power		-106.5		
Electric Power Out			1214.1	
Thermal Power				
Condensate Heat Rejection		1077.0		
Stack Loss		307.9		
Coal Drying		13.0		
Rejected in Solid Waste		10.7		
Internal Electric Power Not Regen.		81.7		
Other Losses to Ambient		128.5		
Thermal Power Out			1618.8	
Total Power Out				2832.9
Unaccounted For				-3.0

- -

Efficiency

Excluding Seed Reprocessing	42.90%
Including Seed Reprocessing	N/A

ENERGY FLOW SUMMARY

Case Number: 1.2

Changes from Reference Case: No O_2 Enrichment

Power In (MW)

Raw Coal to Air Heater Subsystem			1007.7	
Raw Coal to Main Combustor			1875.8	
Total Power In				2883.5

Power Out (MW)

Electric Power				
MHD Generator		590.3		
Steam Turbines				
Shaft Power	858.7			
Main Compressor	-159.9			
Generator Loss	- 15.0			
Net Steam Power		684.8		
Plant Internal Power				
Inverters	- 8.9			
O_2 Plant	0			
Electrical Auxiliaries	- 63.0			
Internal Power		-71.9		
Electric Power Out			1203.2	
Thermal Power				
Condensate Heat Rejection		1138.1		
Stack Loss		300.0		
Coal Drying		44.4		
Rejected in Solid Waste		10.3		
Internal Electric Power Not Regen.		49.5		
Other Losses to Ambient		130.9		
Thermal Power Out			1673.2	
Total Power Out				2876.4
Unaccounted For				7.1

Efficiency

Excluding Seed Reprocessing	41.73%
Including Seed Reprocessing	N/A

ENERGY FLOW SUMMARY

Case Number: 1.3

Changes from Reference Case: 7 Tesla Magnet

Power In (MW)

Raw Coal to Air Heater Subsystem			838.5	
Raw Coal to Main Combustor			2033.8	
Total Power In				2872.3

Power Out (MW)

Electric Power				
MHD Generator		702.7		
Steam Turbines				
Shaft Power	804.3			
Main Compressor	-170.3			
Generator Loss	- 13.6			
Net Steam Power		620.4		
Plant Internal Power				
Inverters	- 10.5			
O_2 Plant	- 32.7			
Electrical Auxiliaries	- 67.6			
Internal Power		-110.8		
Electric Power Out			1212.3	
Thermal Power				
Condensate Heat Rejection		1066.0		
Stack Loss		316.6		
Coal Drying		48.1		
Rejected in Solid Waste		11.1		
Internal Electric Power Not Regen.		85.2		
Other Losses to Ambient		130.4		
Thermal Power Out			1657.4	
Total Power Out				2869.7
Unaccounted For				2.6

Efficiency

Excluding Seed Reprocessing	42.21%
Including Seed Reprocessing	N/A

ENERGY FLOW SUMMARY

Case Number: 1.4

Changes from Reference Case: Single Stage Combustor with 85% Slag Rejection

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Power In (MW)

Raw Coal to Air Heater Subsystem			838.5	
Raw Coal to Main Combustor			2033.8	
Total Power In				2872.3

Power Out (MW)

Electric Power				
MHD Generator		664.4		
Steam Turbines				
Shaft Power	820.4			
Main Compressor	-161.8			
Generator Loss	- 14.1			
Net Steam Power		644.5		
Plant Internal Power				
Inverters	- 10.0			
O_2 Plant	- 37.7			
Electrical Auxiliaries	- 66.7			
Internal Power		109.4		
Electric Power Out			1199.5	
Thermal Power				
Condensate Heat Rejection		1087.3		
Stack Loss		316.6		
Coal Drying		48.1		
Rejected in Solid Waste		11.1		
Internal Electric Power Not Regen.		84.8		
Other Losses to Ambient		130.4		
Thermal Power Out			1678.3	
Total Power Out				2877.8
Unaccounted For				-5.5

Efficiency

Excluding Seed Reprocessing	41.76%
Including Seed Reprocessing	N/A

ENERGY FLOW SUMMARY

Case Number: 1.4a

Changes from Reference Case: Single Stage Combustor with 70% Slag Rejection

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Power In (MW)

Raw Coal to Air Heater Subsystem			838.5	
Raw Coal to Main Combustor			2033.8	
Total Power In				2872.3

Power Out (MW)

Electric Power				
MHD Generator		634.6		
Steam Turbines				
Shaft Power	833.0			
Main Compressor	-152.7			
Generator Loss	- 14.6			
Net Steam Power		665.7		
Plant Internal Power				
Inverters	- 9.5			
O_2 Plant	- 32.7			
Electrical Auxiliaries	- 66.1			
Internal Power		-108.3		
Electric Power Out			1192.0	
Thermal Power				
Condensate Heat Rejection		1104.0		
Stack Loss		316.5		
Coal Drying		48.1		
Rejected in Solid Waste		10.2		
Internal Electric Power Not Regen.		84.5		
Other Losses to Ambient		130.4		
Thermal Power Out			1693.7	
Total Power Out				2885.7
Unaccounted For				-13.4

Efficiency

Excluding Seed Reprocessing	41.50%
Including Seed Reprocessing	N/A

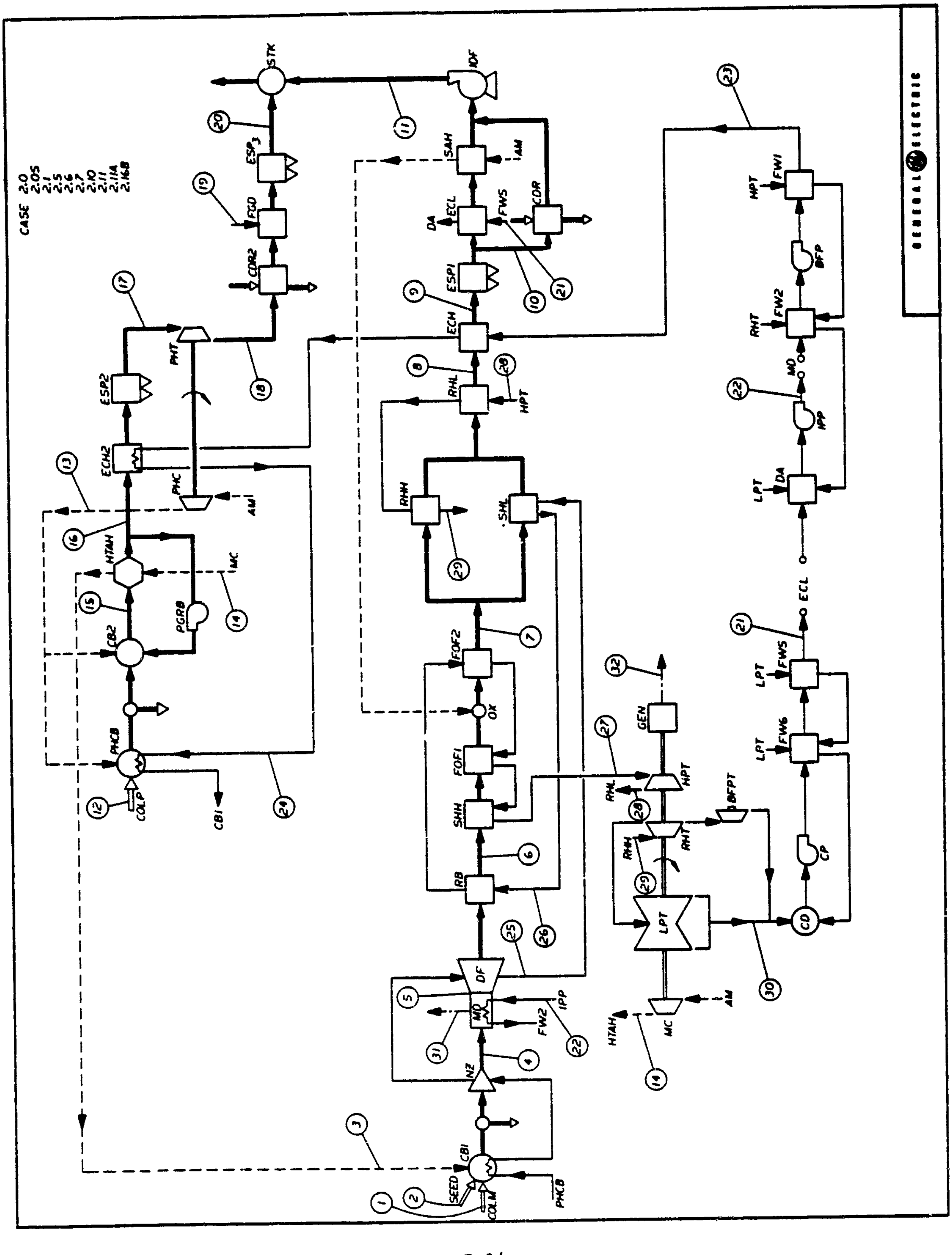
BASE CASE 2

INDEX

Case Number	Overall Efficiency	System Diagram	State Points	Energy Flow Summary
2.0	✓	✓		✓
2.0S	✓	✓		✓
2.0A	✓	✓	✓	✓
2.0B	✓	✓	✓	✓
2.1	✓	✓	✓	✓
2.2	✓	✓		✓
2.2A	✓	✓		✓
2.4	✓	✓		✓
2.4A	✓	✓		✓
2.5	✓	✓		✓
2.6	✓	✓		✓
2.7	✓	✓	✓	✓
2.10	✓	✓		✓
2.11	✓	✓		✓
2.11A	✓	✓		✓
2.12	✓	✓	✓	✓
2.16	✓	✓	✓	✓
2.16A	✓	✓		✓
2.16B	✓	✓		✓
2.17	✓	✓	✓	✓
2.18	✓	✓	✓	✓

Overall Plant Efficiency
Base Case 2

Case No.	Parameter Variation	Efficiency, % Without Seed Reprocessing	With Seed Reprocessing
2.0	Reference System	43.66	43.45
2.0s	70% Slag Rejection	43.12	42.91
2.0a	S^3PMB + Coal 91% Slag Rejection	42.66	42.53
2.0b	S^3PMB + SPMB	43.42	43.22
2.1	Illinois #6 Coal 70% Slag Rejection	44.48	43.80
2.2	2-Stage Cyclone MHD Combustor Hot Bottom HTAH	44.13	43.89
2.2a	2-Stage Cyclone MHD Combustor	43.23	43.02
2.4	2-Stage Cyclone MHD Combustor NASA Generator 20 m Channel	43.47	43.26
2.4a	2-Stage Cyclone MHD Combustor E_y = 4 kV/m Generator 20 m Channel	43.45	43.23
2.5	Cs Seed	44.72	44.57
2.6	Supersonic Generator	43.01	42.81
2.7	8-7T Magnetic Field	44.91	44.69
2.10	1500 MWt MHD Combustor Input	41.98	41.78
2.11	2000 MWt MHD Combustor Input	42.91	42.70
2.11a	2000 MWt MHD Combustor Input 70% Slag Rejection	42.34	42.13
2.12	1 atm. HTAH Reheat Single Stage Preheat Combustor	43.00	42.79
2.16	Hot Bottom HTAH E_y = 4 kV/m Generator	45.49	45.25
2.16a	Hot Bottom HTAH	44.56	44.33
2.16b	E_y = 4 kV/m Channel	44.56	44.34
2.17	CAPFB Preheat Combustor	42.53	42.33
2.18	1200 F Air and 1300 F Recirc. Gas to Preheat Combustor Single Stage Preheat Combustor	44.26	44.03
Best	Illinois #6 Coal 85% Slag Rejection E_y = 4 kVm Generator Hot Bottom HTAH Single Stage Preheat Combustor	47.0 (EST)	46.3 (EST)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IDF
SAH
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CDR
FWS
DA
AM
ESP3
FGD
CDR2
ESP1
ECM
ESP2
PHT
ECH2
PHC
HTAH
PGRB
MC
CB2
PHCB
COLP
CB1
RHL
HPT
RHH
SHL
EOF2
OX
EOF1
SHH
RB
DF
MD
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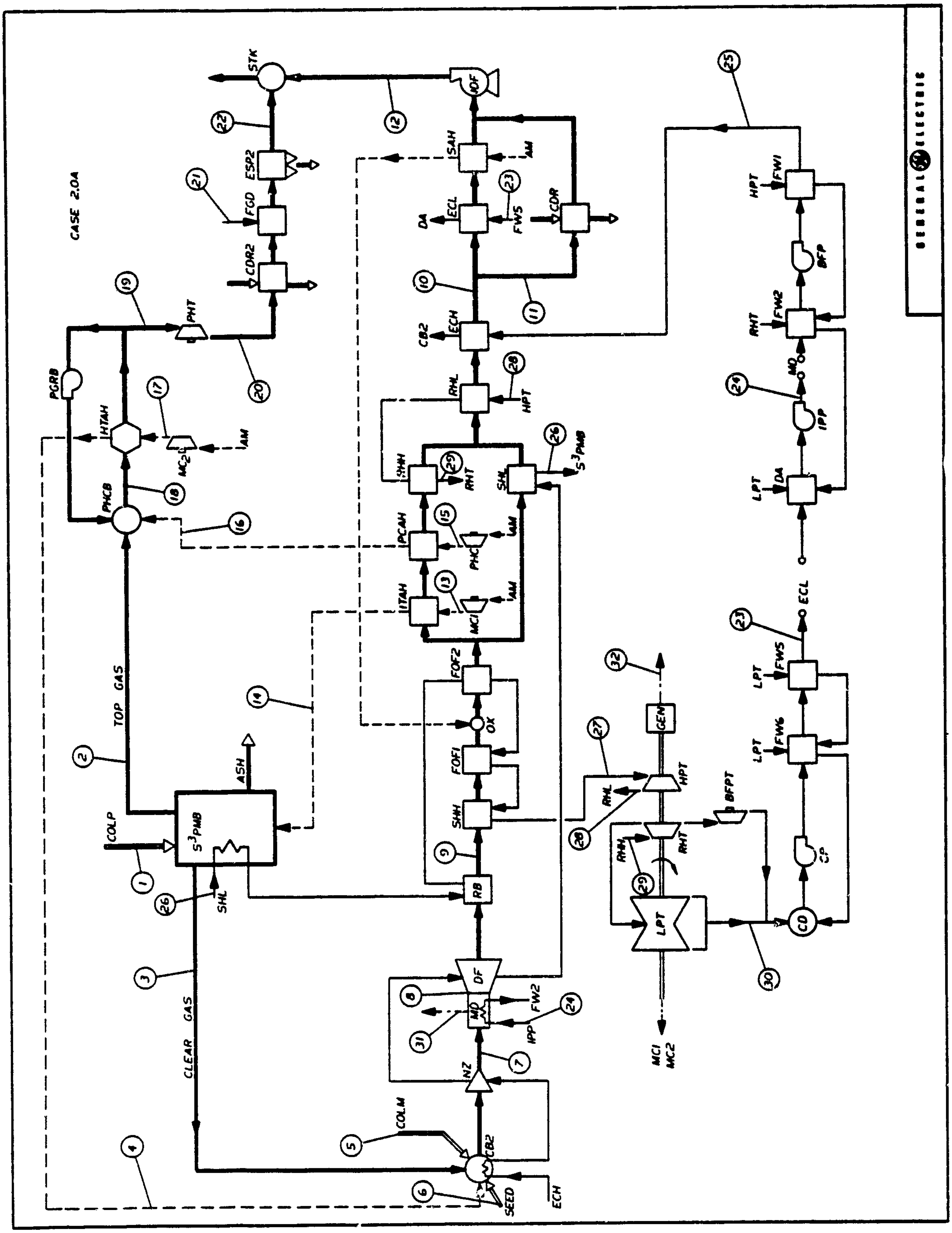
BFPT
CD
CP
FW6
FW1
BFP
HPP
HTAH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HTAH
MC2
PHCB
ECH
CB2
RHL
HPT
RHH
RHT
SHL
S^3PMB
PCAH
PHC
LTAH
MC1
FOF2
OX
FOF1
SHH
RB
DF
MD
FW2
IPP
NZ
COLM
SEED
ECH
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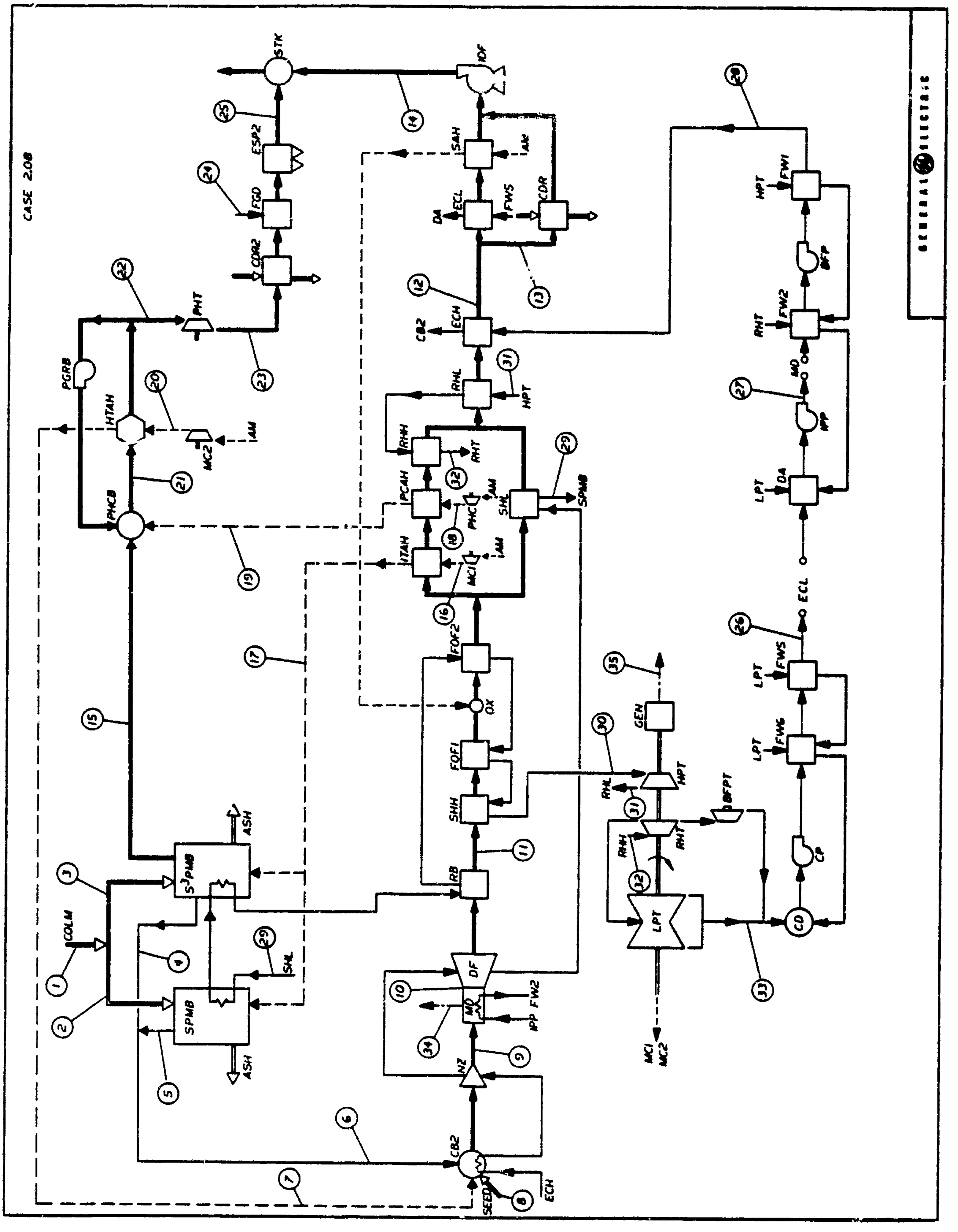
CLEAR GAS
COLP
ASH
SHL
GEN
BFPT
LPT
CD
CP
FW6
FW1
BFP
MC1
MC2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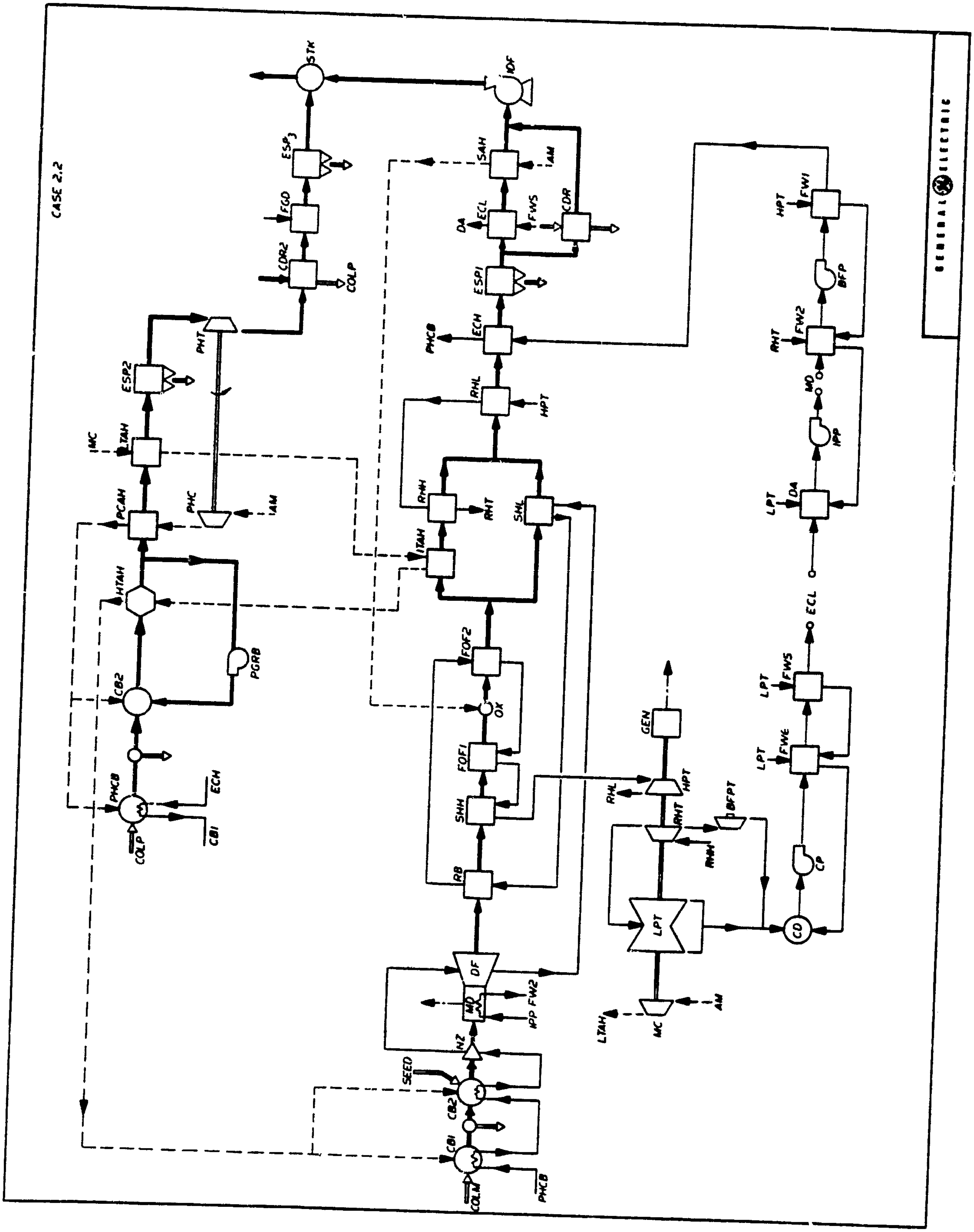

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ECH
PHCB
ESP2
RHL
HPT
IMC
LTAH
PCAH
PHC
IAM
RHH
RHT
SHL
ITAH
HTAH
PGRB
FOF2
CB2
OX
FOF1
SHH
RB
PHCB
ECH
CB1
COLP
GEN
RHL
HPT
RHT
BFPT
RHH
LPT
CD
CP
DF
MD
IPP FW2
NZ
SEED
CB2
CB1
COLM
PHCB
LTAH
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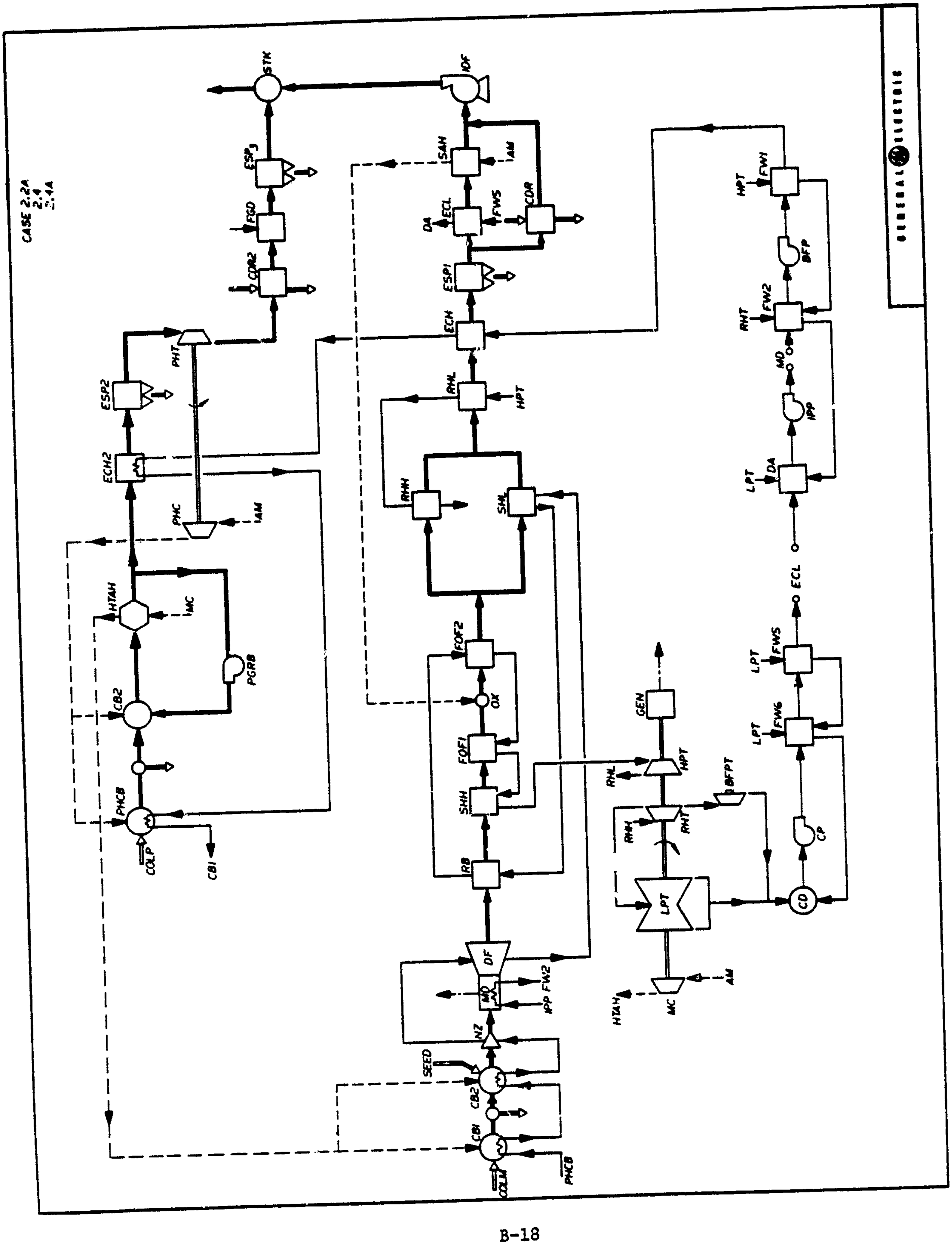
FWE
LPT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2.4
2.4A
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PHC
AM
HTAH
MC
PGRB
CB2
PHCB
COLP
CB1
SAH
ECL
FWS
CDR
DA
ESP1
ECH
RHL
HPT
RHH
SHL
FOF2
OX
FOF1
SHH
RB
DF
MD
IPP
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FW5
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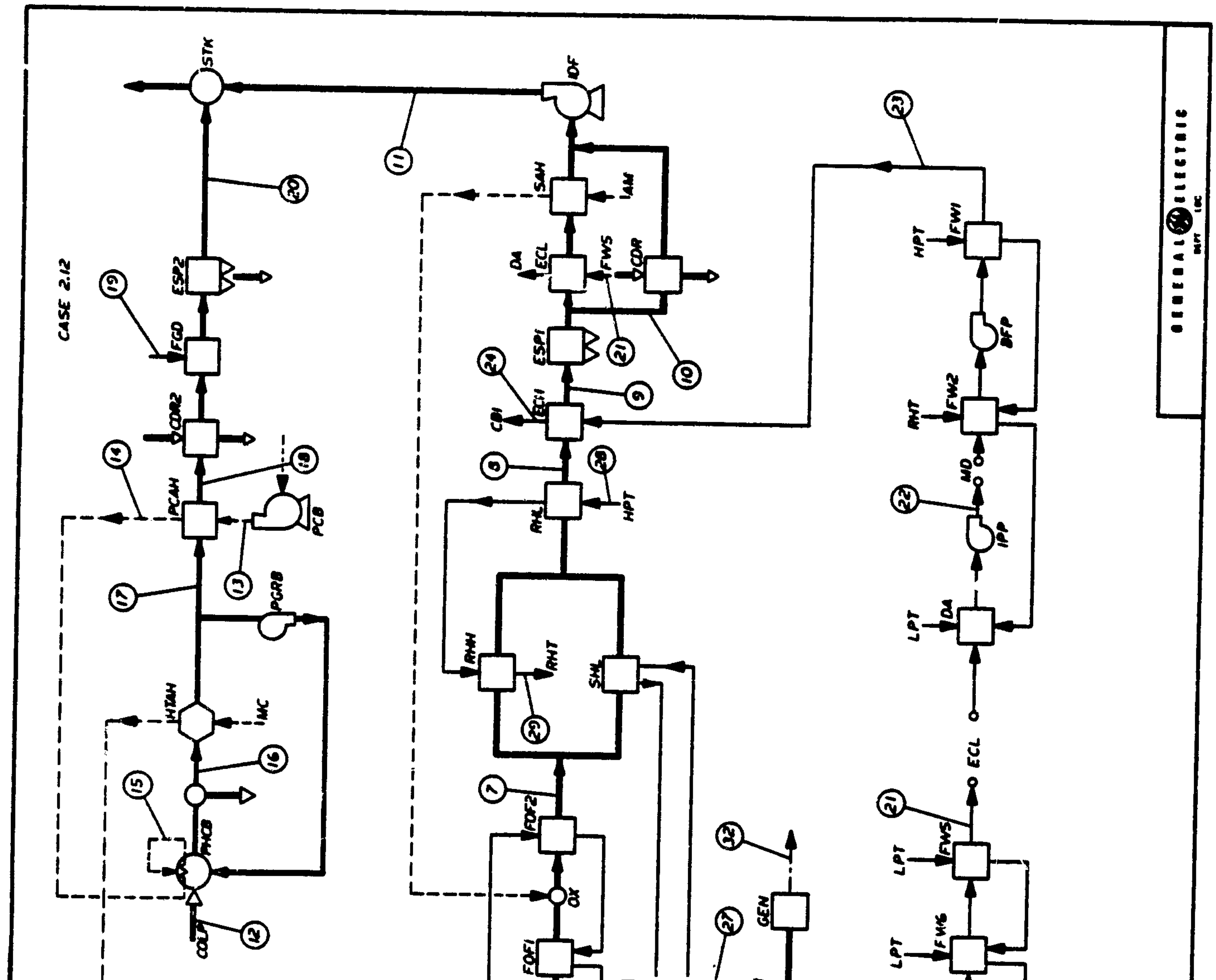
CASE 2.12
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MC
PHCB
COLP
RHL
HPT
CBI
ECH
RHH
RHT
SHL
FOF2
OX
FOF1
SHH
RB
DF
MD
IPP
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BFPT
CD
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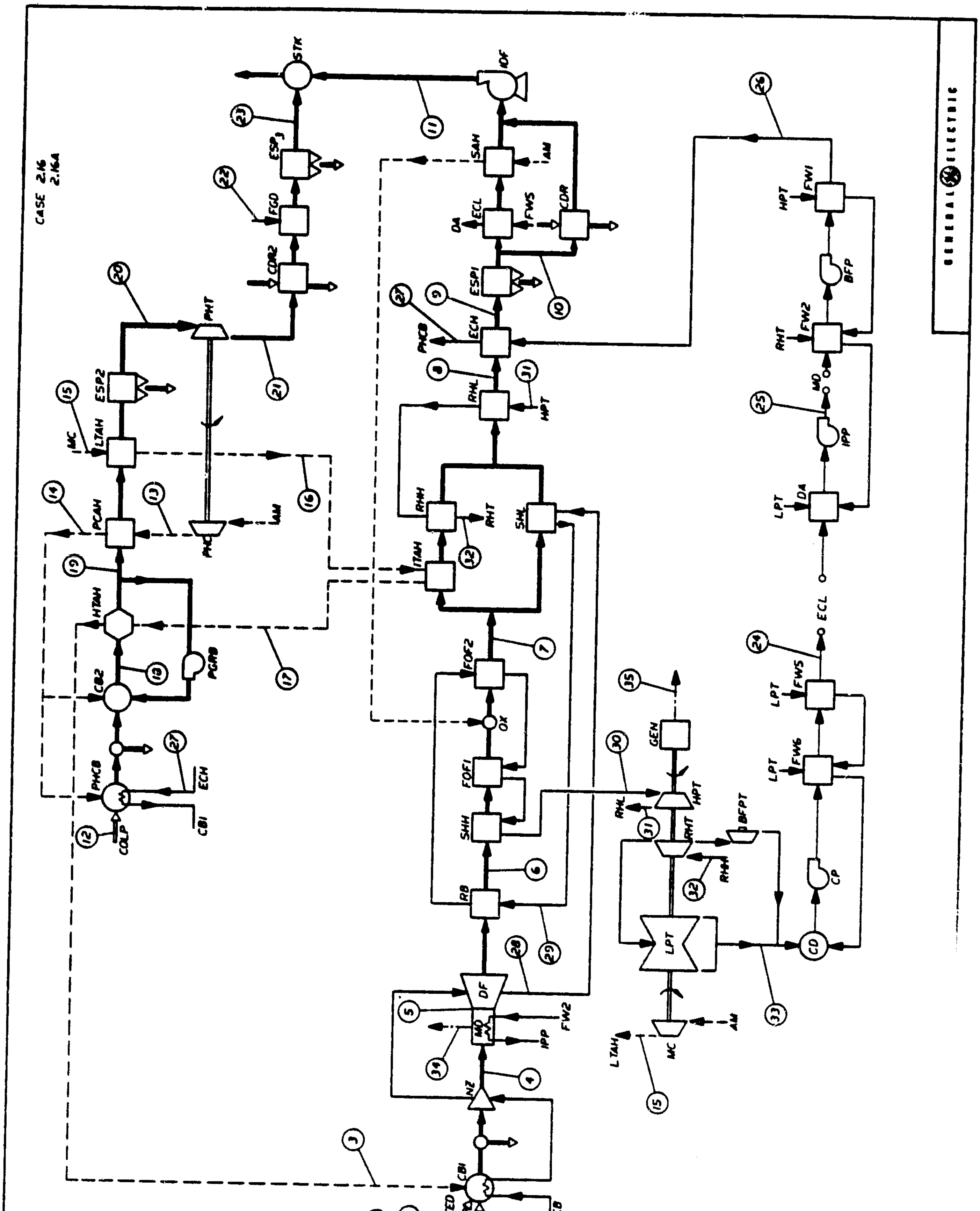
BFP
HT AH
AM
GENERAL ELECTRIC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ESP1
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PHCB
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HPT
RHH
RHT
SHL
ITAH
FOF2
OX
FOF1
SHH
RB
DF
MD
NZ
CBI
SEED
COLM
FW2
IPP
ESP3
FGD
CDR2
PHT
ESP2
LTAH
MC
PCAH
HTAH
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FW1
BFP
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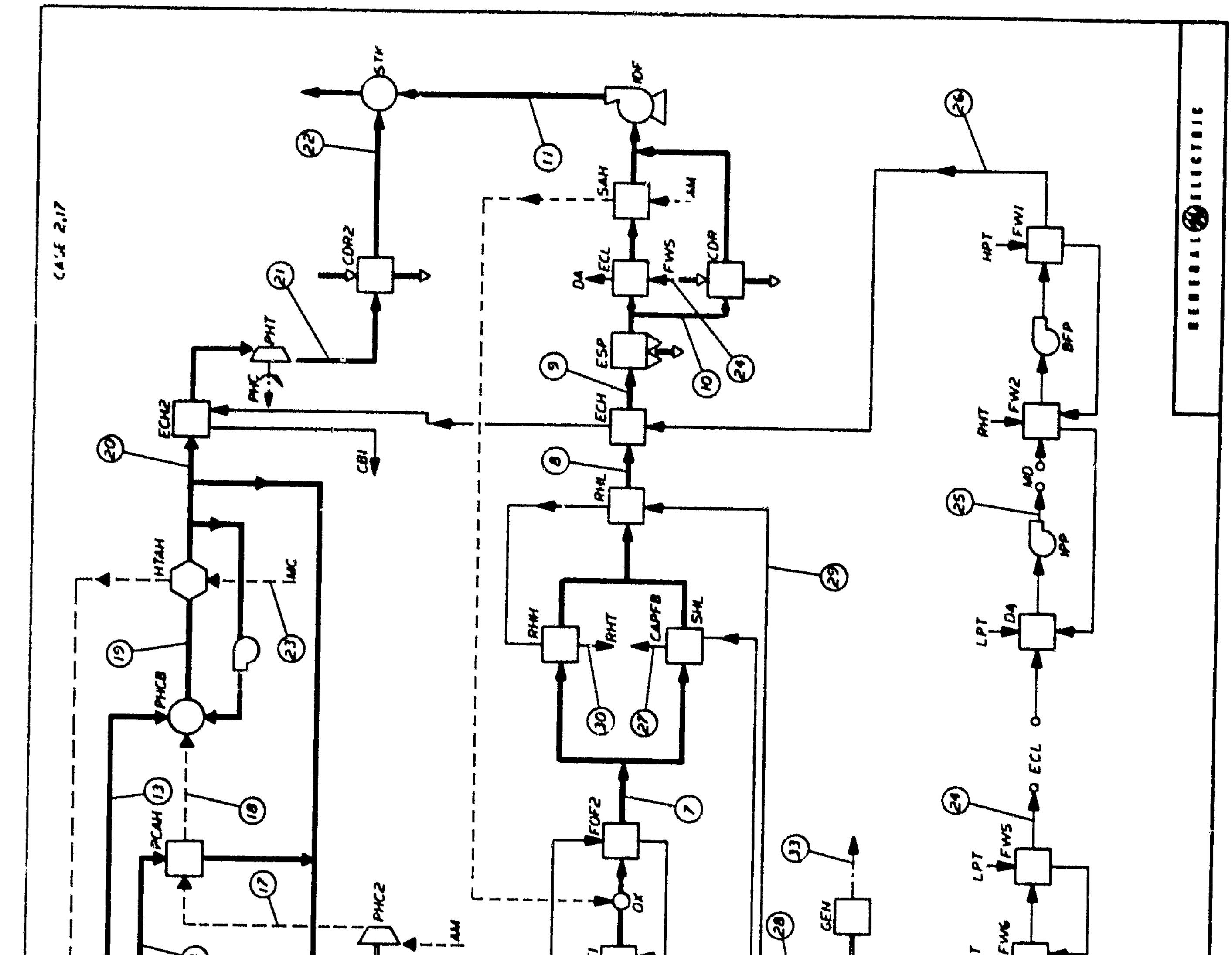
LTAH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CDR
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ESP
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CBI
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RHH
RHT
CAPFB
SHL
HTAH
MC
PHCB
PCAH
FUEL GAS
OFF-GAS
SHL
RB
BED
CAPFB
COLP
PHC1
PHC2
AM
FOF2
FOF1
OX
SHH
RB
GEN
HPT
BFPT
LPT
CD
CP
FW6
FW5
FW2
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COLM
ECH2
HTAH
DA
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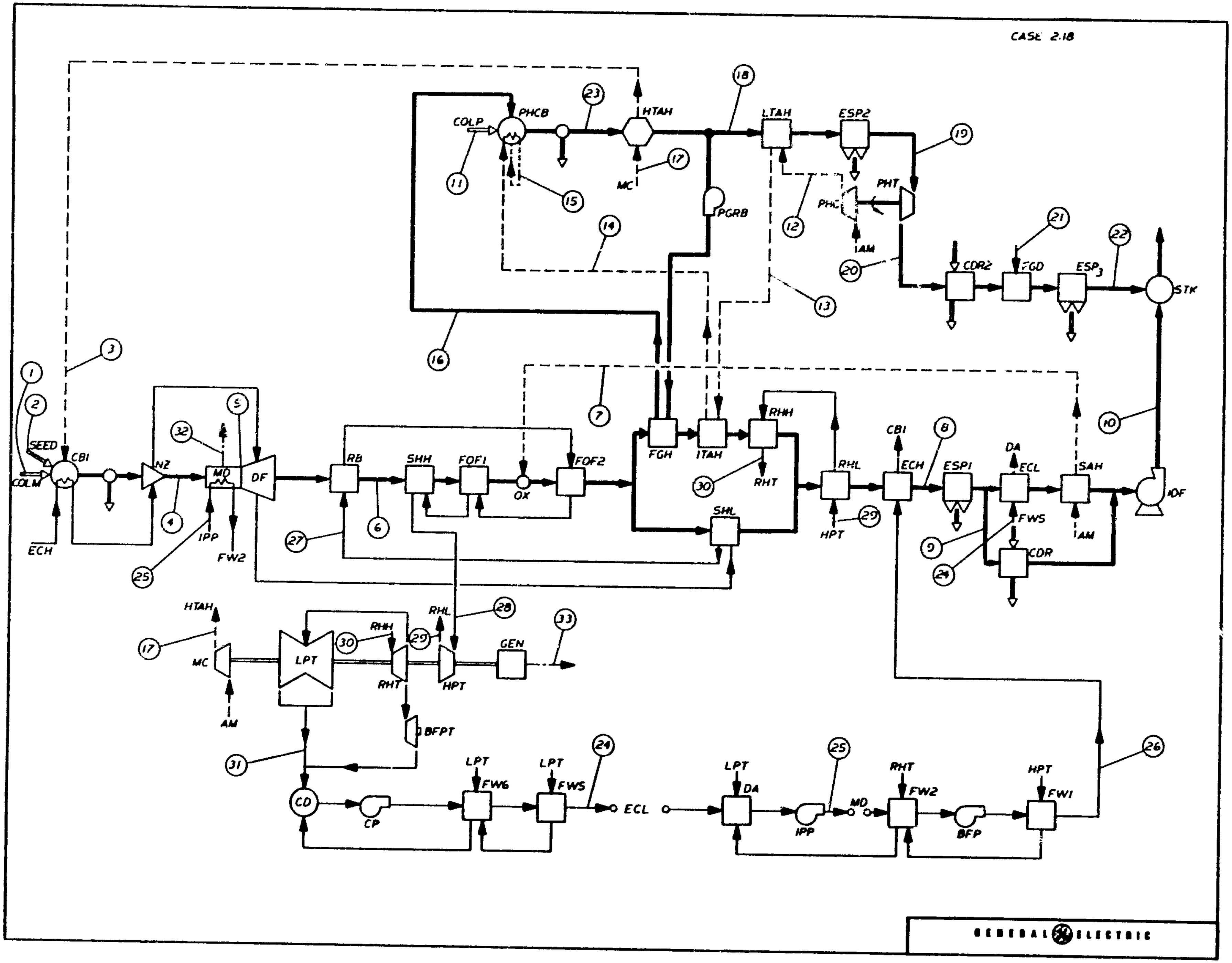
GENERAL ELECTRIC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RHT
SHL
RHL
HPT
ECH
CBI
ESP1
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SAH
AM
IDF
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ESP3
GEN
LPT
BFPT
CD
CP
FW6
FW5
ECL
DA
MD
FW2
BFP
FW1
GENERAL ELECTRIC

State Points, Case 2.0

Location	T (F)/(K)	P (Psia)	m (Kg/s)	E (MW)
1	220/378	-	70.36	1809.2
2	220/378	-	12.7	0.8
3	3000/1922	136.5	529.4	995.6
4	4677/2854	127.0	605.7	2668.4
5	3582/2246	17.6	605.7	1881.3
6	2900/1867	14.55	605.7	1533.7
7	1867/1293	14.44	693.9	875.9
8	952/784	14.27	693.9	439.1
9	611/595	14.17	693.9	288.4
10	611/595	14.17	250.1	104.7
11	260/400	14.75	678.0*	137.1
12	220/378	-	42.19	1084.8
13	648/616	148.5	359.7	121.4
14	623/602	141.9	529.4	170.7
15	3300/2089	136.5	487.8	1167.7
16	985/803	131.1	398.1	269.9
17	877/743	130.7	397.4	241.6
18	373/463	15.15	397.4	117.7
19	59/288	15.15	13.1	0.0
20	162/346	14.75	410.5	101.7
21	190/361	157	564.7	-
22	301/423	400	672.8	-
23	510/534	4230	672.8	-
24	608/593	4200	672.8	-
25	694/641	3950	672.6	-
26	720/656	3850	672.8	-
27	1000/811	3500	672.8	-
28	603/591	768	662.2	-
29	1000/811	691	662.2	-
30	106/314	2.3" Hg	462.7	-
31	-	-	-	658.3
32	-	-	-	661.7

*Excludes coal drying moisture

State Points, Case 2.0a

Location	T (F)/(K)	p (psia)	m (kg/s)	E (MW)
1	220/378	-	68.88	1649.1
2	704/647	101.5	84.7	871.2
3	3340/2111	101.5	150.64	831.6
4	3000/1922	101.5	424.7	794.3
5	220/378	-	42.58	1094.9
6	220/378	-	13.0	0.9
7	4544/2780	98.0	627.1	2641.6
8		17.6	627.1	1919.7
9	2900/1867	14.55	627.1	
10	610/594	14.17	731.5	273.8
11	610/594	14.17	130.4	49.1
12	289/416	14.75	717.3*	132.1
13	548/560	115.0	173.3	46.7
14	1300/978	109.1	173.3	126.1
15	538/554	105.5	233.7	61.5
16	1100/867	101.5	233.7	140.6
17	527/548	105.7	424.7	109.1
18	3300/2089	99.5	370.7	1015.8
19	850/728	95.5	318.4	273.8
20	455/508	15.15	318.4	187.2
21	59/288	15.15	10.2	0.0
22	180/356	14.75	328.6*	157.0
23	190/361	157	516.4	-
24	301/423	400	615.3	-
25	510/539	4230	615.3	-
26	720/656	3850	615.3	-
27	100/811	3500	615.3	-
28	603/591	768	605.6	-
29	1000/811	691	605.6	-
30	106/314	2.3" Hg	423.2	-
31	-	-	-	607.2
32	-	-	-	597.9

*Excludes Coal Drying Moisture

State Points, Case 2.0b

Location	T(F)/(K)	p (Psia)	$\dot{m}$ (kg/s)	$\dot{E}$ (MW)
1	220/378	-	111.64	2673.0
2	220/378	-	61.11	1463.2
3	220/378	-	50.53	1209.8
4	3340/2111	101.5	110.5	610.1
5	2535/1644	101.5	208.2	1511.1
6	2775/1798	101.5	318.7	2121.2
7	3000/1922	101.5	312.1	585.1
8	220/378	-	13.7	0.9
9	4468/2738	98.0	644.5	2635.5
10	-	17.6	644.5	1926.1
11	2900/1867	14.55	644.5	-
12	610/594	14.17	739.2	277.9
13	610/594	14.17	123.9	46.9
14	285/414	14.75	724.9*	133.9
15	704/647	101.5	62.2	639.3
16	555/564	115.6	280.3	76.4
17	1300/978	109.1	280.3	204.0
18	533/552	106.0	171.6	44.8
19	1100/867	103.5	171.6	103.2
20	527/548	105.7	312.1	80.2
21	3300/2089	101.5	272.1	745.6
22	850/728	97.4	233.7	201.0
23	452/507	15.15	233.7	137.0
24	59/288	15.15	7.5	0.0
25	180/356	14.75	241.2*	114.6
26	190/361	157	513.3	-
27	301/423	400	611.6	-
28	510/539	4230	611.6	-
29	720/656	3850	611.6	-
30	1000/811	3500	611.6	-
31	630/591	768	601.9	-
32	1000/811	691	601.9	-
33	106/314	2.3" Hg	420.6	-
34				603.6
35				593.6

*Excludes Coal Drying Moisture

State Points, Case 2.1

Location	T (F)/(K)	p (Psia)	ṁ (kg/s)	E (MW)
1	220/378	-	63.98	1810.9
2	220/378	-	10.7	0.7
3	3000/1922	136.7	528.4	993.7
4	4732/2884	127.2	597.5	2667.0
5	3621/2267	17.7	597.5	1875.0
6	2900/1867	14.65	597.5	1513.0
7	1918/1321	14.54	685.6	866.5
8	957/787	14.37	685.6	413.3
9	611/595	14.27	685.6	262.5
10	611/592	14.27	115.4	44.5
11	282/412	14.75	669.7*	122.5
12	220/378	-	37.42	1059.2
13	649/616	148.7	350.3	118.4
14	623/602	142.1	528.4	170.5
15	3297/2087	136.7	482.8	1147.0
16	955/786	131.3	383.8	248.2
17	894/752	130.9	383.1	232.7
18	382/468	15.15	383.1	111.9
19	59/288	14.75	13.1	0
20	230/383	14.75	396.2*	111.9
21	190/361	157	575.5	-
22	301/423	400	674.8	-
23	510/539	4230	674.8	-
24	601/589	4200	674.8	-
25	689/638	3950	674.8	-
26	720/656	3850	674.8	-
27	1000/811	3500	674.8	-
28	603/591	768	664.4	-
29	1000/811	691	664.4	-
30	106/314	2.3" Hg	472.6	-
31	-	-	-	663.7
32	-	-	-	668.8

*Excludes Coal Drying Moisture

State Points, Case 2.7

Location	T (F)/(K)	p (Psia)	$\dot{m}$ (kg/s)	$\dot{E}$ (MW)
1	220/378	-	70.36	1809.2
2	220/378	-	12.7	0.8
3	3000/1922	137.4	529.4	995.6
4	4678/2854	127.9	605.7	2669.0
5	3472/2184	17.6	605.7	1816.4
6	2900/1867	14.55	605.7	1533.7
7	1804/1258	14.44	693.9	844.9
8	952/784	14.27	693.9	439.1
9	611/595	14.17	696.9	288.4
10	611/595	14.17	250.2	104.7
11	261/401	14.75	678.0*	137.4
12	220/378	-	42.19	1084.8
13	650/617	149.4	359.7	121.9
14	625/603	142.8	529.4	171.4
15	3302/2090	137.4	488.0	1168.2
16	988/804	132.0	398.1	270.8
17	879/743	131.6	397.4	242.0
18	373/463	15.15	397.4	117.6
19	59/288	15.15	13.1	0.0
20	162/346	14.75	410.5*	101.7
21	190/361	157	546.9	-
22	301/423	400	648.3	-
23	510/539	4230	648.3	-
24	612/596	4200	648.3	-
25	699/644	3950	648.3	-
26	720/656	3850	648.3	-
27	1000/811	3500	648.3	-
28	603/591	768	637.7	-
29	1000/811	691	637.7	-
30	106/314	2.3" Hg	448.4	-
31	-	-	-	723.8
32	-	-	-	633.2

*Excludes Coal Drying Moisture

State Points, Case 2.12

Location	T (F)/(K)	p (Psia)	$\dot{m}$ (kg/s)	$\dot{E}$ (MW)
1	220/378	-	70.36	1809.2
2	220/378	-	12.7	0.8
3	3000/1922	136.5	529.4	995.6
4	4677/2854	127.0	605.7	2668.4
5	3582/2246	17.6	605.7	1881.3
6	2900/1867	14.55	605.7	1533.7
7	1940/1333	14.44	693.9	912.7
8	952/784	14.27	693.9	439.1
9	611/595	14.17	693.9	288.4
10	611/595	14.17	96.6	40.4
11	290/417	14.75	678.0*	149.2
12	220/378	-	40.65	1045.2
13	115/319	19.6	346.6	11.2
14	601/19.4	19.4	346.6	107.3
15	1100/867	17.4	346.6	211.8
16	3301/2089	17.05	474.1	1152.1
17	956/787	15.35	383.4	253.0
18	559/566	15.15	383.4	156.5
19	59/288	15.15	12.6	0.0
20	309/427	14.75	396.0*	98.7
21	190/361	157	557.1	-
22	301/423	400	646.1	-
23	510/539	4180	646.1	-
24	597/587	4160	646.1	-
25	672/629	3950	646.1	-
26	720/656	3850	646.1	-
27	1000/811	3500	646.1	-
28	603/591	768	635.5	-
29	1000/811	691	635.5	-
30	106/314	2.3" Hg	458.6	-
31	-	-	-	658.3
32	-	-	-	639.6

*Excludes Coal Drying Moisture

State Points, Case 2.16

Location	T (F)/(K)	p (Psia)	$\dot{m}$ (kg/s)	E (MW)
1	220/378	-	70.36	1809.2
2	220/378	-	12.7	0.8
3	3000/1922	-	529.4	995.6
4	4709/2872	148.9	605.7	2682.3
5	3553/2229	17.6	605.7	1863.9
6	2900/1867	14.55	605.7	1533.7
7	2038/1388	14.44	693.9	962.1
8	953/785	14.27	693.9	439.7
9	612/596	14.17	693.9	289.0
10	612/596	14.17	223.0	93.4
11	265/403	14.75	678.0*	139.1
12	220/378	-	32.99	848.2
13	705/647	172.7	281.3	104.4
14	110/867	170.4	281.3	171.9
15	692/640	171.8	529.4	192.5
16	860/733	169.4	529.4	24.60
17	1300/978	162.9	529.4	390.7
18	3302/2090	158.4	453.6	1085.1
19	1503/1091	153.9	311.4	321.2
20	928/771	153.5	310.9	199.3
21	378/466	15.15	310.9	92.8
22	59/288	15.15	9.9	0.0
23	159/344	14.75	320.8*	77.9
24	190/361	157	504.4	-
25	301/423	400	598.0	-
26	510/539	4220	598.0	-
27	604/591	4200	598.0	-
28	697/643	3950	598.0	-
29	720/656	3850	598.0	-
30	1000/811	3500	598.0	-
31	630/591	768	587.4	-
32	1000/811	691	587.4	-
33	106/314	2.3" Hg	413.7	-
34	-	-	-	715.0
35	-	-	-	549.3

*Excludes Coal Drying Moisture

State Points, Case 2.17

Location	T (F)/(K)	p (Psia)	$\dot{m}$ (kg/s)	$\dot{E}$ (MW)
1	220/378	-	70.36	1809.2
2	220/378	-	12.7	0.8
3	3000/1922	136.5	529.4	995.6
4	4677/2854	127.0	605.7	2668.4
5	3582/2246	17.6	605.7	1881.3
6	2900/1867	14.55	605.7	1533.7
7	1867/1293	14.44	693.9	875.9
8	952/784	14.27	693.9	439.1
9	611/595	14.17	693.9	288.4
10	611/595	14.17	141.4	59.3
11	285/414	14.75	678.0*	146.3
12	220/378	-	51.02	1221.2
13	1650/1172	138.0	200.6	1094.8
14	1600/1144	138.0	61.2	57.3
15	780/689	149.0	81.6	33.1
16	404/480	149.0	134.0	25.0
17	623/602	141.9	270.5	84.4
18	818/710	138.0	270.5	116.9
19	3300/2089	136.5	477.0	1167.3
20	950/783	131.1	447.0	309.5
21	370/461	14.75	447.0	154.8
22	250/394	14.75	447.0*	120.5
23	623/602	141.9	529.4	165.2
24	190/361	157	589.4	-
25	301/423	400	683.6	-
26	510/539	4230	683.6	-
27	720/656	3850	683.6	-
28	1000/811	3500	683.6	-
29	603/591	768	672.3	-
30	1000/811	691	672.3	-
31	106/314	2.3" Hg	485.2	-
32	-	-	-	658.3
33	-	-	-	686.6

*Excludes Coal Drying Moisture

State Points, Case 2.18

Location	T (F)/(K)	p (Psia)	ṁ (kg/s)	E (MW)
1	220/378	-	70.36	1809.2
2	220/378	-	12.7	0.8
3	3000/1922	136.5	529.4	995.6
4	4677/2854	127.0	605.7	2668.4
5	3582/2246	17.6	605.7	1881.3
6	2900/1865	14.55	605.7	1533.7
7	200/367	14.47	88.2	7.1
8	611/595	14.17	693.9	288.4
9	611/595	14.17	217.2	90.9
10	267/	14.75	678.0*	139.4
11	220/378	-	33.89	871.4
12	661/623	154.0	289.0	101.8
13	678/632	153.0	289.0	107.0
14	1202/923	151.7	289.0	192.9
15	1300/978	138.5	289.0	210.3
16	1300/978	138.5	162.3	144.7
17	623/602	141.9	529.4	170.7
18	950/783	131.1	318.9	208.9
19	926/770	130.8	318.5	203.7
20	403/479	15.15	318.5	99.0
21	59/288	15.15	10.2	0.0
22	180/356	14.75	328.7*	81.6
23	3300/2089	136.5	481.2	1152.3
24	190/361	157	496.4	-
25	301/423	400	610.9	-
26	510/539	4230	610.9	-
27	720/656	3850	610.9	-
28	1000/811	3500	610.9	-
29	603/811	768	601.3	-
30	1000/811	691	601.3	-
31	106/314	2.3" Hg	420.1	-
32	-	-	-	658.3
33	-	-	-	585.0

*Excludes Coal Drying Moisture

ENERGY FLOW SUMMARY

Case Number: 2.0

Changes from Reference Case: Reference Case

- -

Power In (MW)				
Raw Coal to Air Heater Subsystem			1079.5	
Raw Coal to Main Combustor			1800.2	
Total Power In				2879.7
Power Out (MW)				
Electric Power				
MHD Generator		658.3		
Steam Turbines				
Shaft Power	843.0			
Main Compressor	-171.6			
Generator Loss	- 9.7			
Net Steam Power		661.7		
Plant Internal Power				
Inverters	9.9			
O_2 Plant	0.0			
Electrical Auxiliaries	52.7			
Internal Power		-62.6		
Electric Power Out			1257.4	
Thermal Power				
Condensate Heat Rejection		1136.7		
Stack Loss		238.8		
Coal Drying		68.4		
Rejected in Solid Waste		14.9		
Internal Electric Power Not Regen.		48.5		
Other Losses to Ambient		105.2		
Thermal Power Out			1612.5	
Total Power Out				2869.9
Unaccounted For				9.8
Efficiency				
Excluding Seed Reprocessing	43.66%			
Including Seed Reprocessing	43.45%			

ENERGY FLOW SUMMARY

Case Number: 2.0S

Changes from Reference Case: 70% Slag Rejection

Power In (MW)

Raw Coal to Air Heater Subsystem			1094.2	
Raw Coal to Main Combustor			1800.2	
Total Power In				2894.4

Power Out (MW)

Electric Power				
MHD Generator		623.0		
Steam Turbines				
Shaft Power	858.0			
Main Compressor	-160.3			
Generator Loss	- 10.2			
Net Steam Power		687.5		
Plant Internal Power				
Inverters	9.3			
O_2 Plant	0.0			
Electrical Auxiliaries	53.0			
Internal Power		-62.3		
Electric Power Out			1248.2	
Thermal Power				
Condensate Heat Rejection		1156.9		
Stack Loss		240.2		
Coal Drying		68.7		
Rejected in Solid Waste		15.1		
Internal Electric Power Not Regen.		48.8		
Other Losses to Ambient		106.0		
Thermal Power Out			1635.7	
Total Power Out				2883.9
Unaccounted For				10.5

Efficiency

Excluding Seed Reprocessing	43.12%
Including Seed Reprocessing	42.91%

ENERGY FLOW SUMMARY

Case Number: 2.0a

Changes from Reference Case: S^3PMB Gasifier + Coal to MHD Combustor

- -

Power In (MW)

Raw Coal to S^3PMB			1641.4	
Raw Coal to Main Combustor			1090.1	
Total Power In				2731.5

- -

Power Out (MW)

Electric Power				
MHD Generator		607.2		
Steam Turbines				
Shaft Power	769.4			
Main Compressor	-162.6			
Generator Loss	- 8.9			
Net Steam Power		597.9		
Plant Internal Power				
Inverters	9.1			
O_2 Plant	0.0			
Electrical Auxiliaries	30.6			
Internal Power		-39.7		
Electric Power Out			1165.4	
Thermal Power				
Condensate Heat Rejection		1049.5		
Stack Loss		290.3		
Coal Drying		51.6		
Rejected in Solid Waste		17.5		
Internal Electric Power Not Regen.		44.3		
Other Losses to Ambient		110.2		
Thermal Power Out			1563.4	
Total Power Out				2728.8
Unaccounted For				2.7

- -

Efficiency

Excluding Seed Reprocessing	42.66%
Including Seed Reprocessing	42.53%

ENERGY FLOW SUMMARY

Case Number: 2.0b

Changes from Reference Case: S^3PMB + SPMB Gasifiers for 1st Stage of MHD & Preheat Combustors

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Power In (MW)

Raw Coal to S^3PMB			1202.1	
Raw Coal to SPMB			1453.8	
Total Power In				2655.9

Power Out (MW)

Electric Power				
MHD Generator		603.6		
Steam Turbines				
Shaft Power	765.8			
Main Compressor	-163.4			
Generator Loss	- 8.8			
Net Steam Power		593.6		
Plant Internal Power				
Inverters	9.1			
O_2 Plant	0.0			
Electrical Auxiliaries	35.3			
Internal Power		-44.4		
Electric Power Out			1152.8	
Thermal Power				
Condensate Heat Rejection		1040.2		
Stack Loss		248.4		
Coal Drying		42.1		
Rejected in Solid Waste		18.5		
Internal Electric Power Not Regen.		45.4		
Other Losses to Ambient		101.9		
Thermal Power Out			1496.5	
Total Power Out				2649.3
Unaccounted For				6.6

Efficiency

Excluding Seed Reprocessing	43.42%
Including Seed Reprocessing	43.22%

ENERGY FLOW SUMMARY

Case Number: 2.1

Changes from Reference Case: Illinois #6 Coal, 70% Slag Rejection in Main Combustor

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Power In (MW)

Raw Coal to Air Heater Subsystem			1055.4	
Raw Coal to Main Combustor			1804.6	
Total Power In				2860.0

Power Out (MW)

Electric Power				
MHD Generator		663.7		
Steam Turbines				
Shaft Power	850.1			
Main Compressor	-171.4			
Generator Loss	- 9.9			
Net Steam Power		668.8		
Plant Internal Power				
Inverters	10.0			
O_2 Plant	0.0			
Electrical Auxiliaries	50.5			
Internal Power		-60.5		
Electric Power Out			1272.0	
Thermal Power				
Condensate Heat Rejection		1158.2		
Stack Loss		234.4		
Coal Drying		19.9		
Rejected in Solid Waste		15.1		
Internal Electric Power Not Regen.		47.7		
Other Losses to Ambient		106.4		
Thermal Power Out			1581.7	
Total Power Out				2853.7
Unaccounted For				6.3

Efficiency

Excluding Seed Reprocessing	44.48%
Including Seed Reprocessing	43.80%

ENERGY FLOW SUMMARY

Case Number: 2.2

Changes from Reference Case: 2 Stage Cyclone Main Combustor, Hot Bottom HTAH (1300 F Air In)

Power In (MW)

Raw Coal to Air Heater Subsystem			844.1	
Raw Coal to Main Combustor			1800.2	
Total Power In				2644.3

Power Out (MW)

Electric Power				
MHD Generator		636.7		
Steam Turbines				
Shaft Power	776.9			
Main Compressor	-176.3			
Generator Loss	8.8			
Net Steam Power		591.6		
Plant Internal Power				
Inverters	9.6			
O_2 Plant	0.0			
Electrical Auxiliaries	51.8			
Internal Power		-61.4		
Electric Power Out			1166.9	
Thermal Power				
Condensate Heat Rejection		1048.2		
Stack Loss		217.1		
Coal Drying		61.3		
Rejected in Solid Waste		17.8		
Internal Electric Power Not Regen.		47.0		
Other Losses to Ambient		95.1		
Thermal Power Out			1486.5	
Total Power Out				2653.4
Unaccounted For				-9.1

Efficiency

Excluding Seed Reprocessing	44.13%
Including Seed Reprocessing	43.89%

ENERGY FLOW SUMMARY

Case Number: 2.2A

Changes from Reference Case: Two Stage Cyclone Main Combustor

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Power In (MW)

Raw Coal to Air Heater Subsystem			1083.4	
Raw Coal to Main Combustor			1800.2	
Total Power In				2883.6

Power Out (MW)

Electric Power				
MHD Generator		636.7		
Steam Turbines				
Shaft Power	851.5			
Main Compressor	-169.3			
Generator Loss	- 10.0			
Net Steam Power		672.2		
Plant Internal Power				
Inverters	9.6			
O_2 Plant	0.0			
Electrical Auxiliaries	52.7			
Internal Power		-62.3		
Electric Power Out			1246.6	
Thermal Power				
Condensate Heat Rejection		1149.1		
Stack Loss		239.2		
Coal Drying		68.5		
Rejected in Solid Waste		14.9		
Internal Electric Power Not Regen.		48.2		
Other Losses to Ambient		104.8		
Thermal Power Out			1624.7	
Total Power Out				2871.3
Unaccounted For				12.3

Efficiency

Excluding Seed Reprocessing	43.23%
Including Seed Reprocessing	43.02%

ENERGY FLOW SUMMARY

Case Number: 2.4

Changes from Reference Case: NASA Generator, L=20m, Optimized. 2 Stage Cyclone MHD Combustor

Power In (MW)

Raw Coal to Air Heater Subsystem			1057.3	
Raw Coal to Main Combustor			1800.2	
Total Power In				2857.5

Power Out (MW)

Electric Power				
MHD Generator		643.0		
Steam Turbines				
Shaft Power	848.9			
Main Compressor	-188.9			
Generator Loss	- 9.6			
Net Steam Power		650.4		
Plant Internal Power				
Inverters	9.6			
O_2 Plant	0.0			
Electrical Auxiliaries	41.5			
Internal Power		-51.1		
Electric Power Out			1242.3	
Thermal Power				
Condensate Heat Rejection		1144.7		
Stack Loss		225.7		
Coal Drying		67.9		
Rejected in Solid Waste		14.8		
Internal Electric Power Not Regen.		48.0		
Other Losses to Ambient		104.7		
Thermal Power Out			1605.8	
Total Power Out				2848.1
Unaccounted For				9.4

Efficiency

Excluding Seed Reprocessing	43.47%
Including Seed Reprocessing	43.26%

ENERGY FLOW SUMMARY

Case Number: 2.4a

Changes from Reference Case: GE Generator, L=20m Ey=4 kV/m
2 Stage Cyclone MHD Combustor

Power In (MW)

Raw Coal to Air Heater Subsystem			1090.0	
Raw Coal to Main Combustor			1800.2	
Total Power In				2890.2

Power Out (MW)

Electric Power				
MHD Generator		641.7		
Steam Turbines				
Shaft Power	850.3			
Main Compressor	-163.5			
Generator Loss	- 10.0			
Net Steam Power		676.8		
Plant Internal Power				
Inverters	9.6			
O_2 Plant	0.0			
Electrical Auxiliaries	53.2			
Internal Power		-62.8	1255.7	
Electric Power Out				
Thermal Power				
Condensate Heat Rejection		1146.6		
Stack Loss		239.8		
Coal Drying		68.6		
Rejected in Solid Waste		15.0		
Internal Electric Power Not Regen.		48.7		
Other Losses to Ambient		106.2		
Thermal Power Out			1624.9	
Total Power Out				2880.6
Unaccounted For				9.6

Efficiency

Excluding Seed Reprocessing	43.45%
Including Seed Reprocessing	43.23%

ENERGY FLOW SUMMARY

Case Number: 2.5

Changes from Reference Case: Cesium Seed

- -

Power In (MW)				
Raw Coal to Air Heater Subsystem			1063.8	
Raw Coal to Main Combustor			1800.2	
Total Power In				2864.0
Power Out (MW)				
Electric Power				
MHD Generator		720.6		
Steam Turbines				
Shaft Power	816.2			
Main Compressor	-184.0			
Generator Loss	- 9.5			
Net Steam Power		622.7		
Plant Internal Power				
Inverters	10.8			
O_2 Plant	0.0			
Electrical Auxiliaries	51.7			
Internal Power		-62.5		
Electric Power Out			1280.8	
Thermal Power				
Condensate Heat Rejection		1100.6		
Stack Loss		237.3		
Coal Drying		68.0		
Rejected in Solid Waste		14.8		
Internal Electric Power Not Regen		48.4		
Other Losses to Ambient		103.7		
Thermal Power Out			1572.8	
Total Power Out				2853.6
Unaccounted For				10.4

- -

Efficiency

Excluding Seed Reprocessing	44.72%
Including Seed Reprocessing	44.57%

ENERGY FLOW SUMMARY

Case Number: 2.6

Changes from Reference Case: Supersonic Channel

- -

Power In (MW)

Raw Coal to Air Heater Subsystem			1118.8	
Raw Coal to Main Combustor			1800.2	
Total Power In				2919.0

- -

Power Out (MW)

Electric Power				
MHD Generator		605.1		
Steam Turbines				
Shaft Power	866.7			
Main Compressor	-142.2			
Generator Loss	- 10.6			
Net Steam Power		713.9		
Plant Internal Power				
Inverters	9.1			
O_2 Plant	0.0			
Electrical Auxiliaries	54.3			
Internal Power		-63.4		
Electric Power Out			1255.6	
Thermal Power				
Condensate Heat Rejection		1168.6		
Stack Loss		242.5		
Coal Drying		69.3		
Rejected in Solid Waste		15.1		
Internal Electric Power Not Regen.		49.3		
Other Losses to Ambient		108.6		
Thermal Power Out			1653.4	
Total Power Out				2909.0
Unaccounted For				10.0

- -

Efficiency

Excluding Seed Reprocessing	43.01%
Including Seed Reprocessing	42.81%

ENERGY FLOW SUMMARY

Case Number: 2.7

Changes from Reference Case: 8T Tapered to 7T Magnetic Field

- -

Power In (MW)

Raw Coal to Air Heater Subsystem			1079.5	
Raw Coal to Main Combustor			1800.2	
Total Power In				2879.7

- -

Power Out (MW)

Electric Power				
MHD Generator		723.8		
Steam Turbines				
Shaft Power	814.9			
Main Compressor	-172.3			
Generator Loss	- 9.4			
Net Steam Power		633.2		
Plant Internal Power				
Inverters	10.9			
O_2 Plant	0.0			
Electrical Auxiliaries	52.9			
Internal Power		-63.8		
Electric Power Out			1293.2	
Thermal Power				
Condensate Heat Rejection		1099.6		
Stack Loss		239.0		
Coal Drying		68.4		
Rejected in Solid Waste		14.9		
Internal Electric Power Not Regen.		48.4		
Other Losses to Ambient		105.2		
Thermal Power Out			1575.5	
Total Power Out				2868.7
Unaccounted For				11.0

- -

Efficiency

Excluding Seed Reprocessing	44.91%
Including Seed Reprocessing	44.69%

ENERGY FLOW SUMMARY

Case Number: 2.10

Changes from Reference Case: 1500 MWt MHD Combustor Input

Power In (MW)

Raw Coal to Air Heater Subsystem			594.5	
Raw Coal to Main Combustor			964.4	
Total Power In				1558.9

Power Out (MW)

Electric Power		298.6		
MHD Generator				
Steam Turbines				
Shaft Power	474.9			
Main Compressor	-79.7			
Generator Loss	- 5.8			
Net Steam Power		389.4		
Plant Internal Power				
Inverters	4.5			
O_2 Plant	0.0			
Electrical Auxiliaries	29.0			
Internal Power		-33.5		
Electric Power Out			654.5	
Thermal Power				
Condensate Heat Rejection		640.5		
Stack Loss		129.5		
Coal Drying		37.0		
Rejected in Solid Waste		8.1		
Internal Electric Power Not Regen.		25.9		
Other Losses to Ambient		58.1		
Thermal Power Out			899.1	
Total Power Out				1553.6
Unaccounted For				5.3

Efficiency

Excluding Seed Reprocessing	41.98%
Including Seed Reprocessing	41.78%

ENERGY FLOW SUMMARY

Case Number: 2.11

Changes from Reference Case: 2000 MWt MHD Combustor Input

Power In (MW)

Raw Coal to Air Heater Subsystem			781.6	
Raw Coal to Main Combustor			1285.9	
Total Power In				2067.5

Power Out (MW)

Electric Power				
MHD Generator		437.4		
Steam Turbines				
Shaft Power	616.4			
Main Compressor	-114.4			
Generator Loss	- 7.3			
Net Steam Power		494.7		
Plant Internal Power				
Inverters	6.6			
O_2 Plant	0.0			
Electrical Auxiliaries	38.3			
Internal Power		-44.9		
Electric Power Out			887.2	
Thermal Power				
Condensate Heat Rejection		831.1		
Stack Loss		171.6		
Coal Drying		49.1		
Rejected in Solid Waste		10.7		
Internal Electric Power Not Regen.		34.8		
Other Losses to Ambient		76.2		
Thermal Power Out			1173.5	
Total Power Out				2060.7
Unaccounted For				6.8

Efficiency

Excluding Seed Reprocessing	42.91%
Including Seed Reprocessing	42.70%

ENERGY FLOW SUMMARY

Case Number: 2.11a

Changes from Reference Case: 2000 MWt MHD Combustor Input
70% Slag Rejection

Power In (MW)

Raw Coal to Air Heater Subsystem			794.3	
Raw Coal to Main Combustor			1285.9	
Total Power In				2080.2

Power Out (MW)

Electric Power				
MHD Generator		410.5		
Steam Turbines				
Shaft Power	628.1			
Main Compressor	-105.1			
Generator Loss	- 7.6			
Net Steam Power		515.4		
Plant Internal Power				
Inverters	6.2			
O_2 Plant	0.0			
Electrical Auxiliaries	39.0			
Internal Power		-45.2		
Electric Power Out			880.7	
Thermal Power				
Condensate Heat Rejection		846.9		
Stack Loss		172.8		
Coal Drying		49.4		
Rejected in Solid Waste		10.8		
Internal Electric Power Not Regen.		35.1		
Other Losses to Ambient		77.5		
Thermal Power Out			1192.5	
Total Power Out				2073.2
Unaccounted For				7.0

Efficiency

Excluding Seed Reprocessing	42.34%
Including Seed Reprocessing	42.13%

ENERGY FLOW SUMMARY

Case Number: 2.12

Changes from Reference Case: Atmospheric, Regeneratively Air-Cooled, Preheat Combustor

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Power In (MW)

Raw Coal to Air Heater Subsystem			1040.1	
Raw Coal to Main Combustor			1800.2	
Total Power In				2840.3

Power Out (MW)

Electric Power				
MHD Generator		658.3		
Steam Turbines				
Shaft Power	818.6			
Main Compressor	-169.5			
Generator Loss	- 9.5			
Net Steam Power		639.6		
Plant Internal Power				
Inverters	9.9			
O_2 Plant	0.0			
Electrical Auxiliaries	66.6			
Internal Power		-76.5		
Electric Power Out			1221.4	
Thermal Power				
Condensate Heat Rejection		1119.8		
Stack Loss		247.9		
Coal Drying		69.2		
Rejected in Solid Waste		13.8		
Internal Electric Power Not Regen.		46.4		
Other Losses to Ambient		115.2		
Thermal Power Out			1612.3	
Total Power Out				2833.7
Unaccounted For				6.6

Efficiency

Excluding Seed Reprocessing	43.00%
Including Seed Reprocessing	42.79%

ENERGY FLOW SUMMARY

Case Number: 2.16

Changes from Reference Case: Hot Bottom HTAH (1300 F Air In)
E_y= 4 kV/m Channel

Power In (MW)

Raw Coal to Air Heater Subsystem			844.1	
Raw Coal to Main Combustor			1800.2	
Total Power In				2644.3

Power Out (MW)

Electric Power				
MHD Generator		715.0		
Steam Turbines				
Shaft Power	750.8			
Main Compressor	-193.4			
Generator Loss	- 8.1			
Net Steam Power		549.3		
Plant Internal Power				
Inverters	10.7			
O_2 Plant	0.0			
Electrical Auxiliaries	50.7			
Internal Power		-61.4		
Electric Power Out			1202.9	
Thermal Power				
Condensate Heat Rejection		1013.3		
Stack Loss		217.1		
Coal Drying		61.3		
Rejected in Solid Waste		17.8		
Internal Electric Power Not Regen.		47.0		
Other Losses to Ambient		93.2		
Thermal Power Out			1449.7	
Total Power Out				2652.6
Unaccounted For				-8.3

Efficiency

Excluding Seed Reprocessing	45.49%
Including Seed Reprocessing	45.25%

ENERGY FLOW SUMMARY

Case Number: 2.16a

Changes from Reference Case: Hot Bottom HTAH (1300 F Air In)

Power In (MW)				
Raw Coal to Air Heater Subsystem			844.1	
Raw Coal to Main Combustor			1800.2	
Total Power In				2644.3
Power Out (MW)				
Electric Power				
MHD Generator		658.3		
Steam Turbines				
Shaft Power	768.0			
Main Compressor	-178.1			
Generator Loss	- 8.4			
Net Steam Power		581.5		
Plant Internal Power				
Inverters	9.9			
O_2 Plant	0.0			
Electrical Auxiliaries	51.5			
Internal Power		-61.4		
Electric Power Out			1178.4	
Thermal Power				
Condensate Heat Rejection		1036.3		
Stack Loss		217.1		
Coal Drying		61.3		
Rejected in Solid Waste		17.8		
Internal Electric Power Not Regen.		47.0		
Other Losses to Ambient		93.6		
Thermal Power Out			1473.1	
Total Power Out				2651.5
Unaccounted For				-7.2

Efficiency

Excluding Seed Reprocessing	44.56%
Including Seed Reprocessing	44.33%

ENERGY FLOW SUMMARY

Case Number: 2.16b

Changes from Reference Case: Ey = 4 kV/m Channel

Power In (MW)

Raw Coal to Air Heater Subsystem			1057.3	
Raw Coal to Main Combustor			1800.2	
Total Power In				2857.5

Power Out (MW)

Electric Power				
MHD Generator		715.0		
Steam Turbines				
Shaft Power	818.4			
Main Compressor	-188.7			
Generator Loss	- 9.2			
Net Steam Power		620.5		
Plant Internal Power				
Inverters	10.7			
O_2 Plant	0.0			
Electrical Auxiliaries	51.4			
Internal Power		-62.1		
Electric Power Out			1273.4	
Thermal Power				
Condensate Heat Rejection		1103.5		
Stack Loss		236.7		
Coal Drying		67.9		
Rejected in Solid Waste		14.8		
Internal Electric Power Not Regen.		48.0		
Other Losses to Ambient		103.1		
Thermal Power Out			1574.0	
Total Power Out				2847.4
Unaccounted For				10.1

Efficiency

Excluding Seed Reprocessing	44.56%
Including Seed Reprocessing	44.34%

ENERGY FLOW SUMMARY

Case Number: 2.17

Changes from Reference Case: CAPFB for 1st Stage of Preheat Combustor

- -

Power In (MW)

Raw Coal to Air Heater Subsystem			1213.7	
Raw Coal to Main Combustor			1800.2	
Total Power In				3013.9

- -

Power Out (MW)

Electric Power				
MHD Generator		658.3		
Steam Turbines				
Shaft Power	868.4			
Main Compressor	-171.6			
Generator Loss	- 10.2			
Net Steam Power		686.6		
Plant Internal Power				
Inverters	9.9			
O_2 Plant	0.0			
Electrical Auxiliaries	53.2			
Internal Power		-63.1		
Electric Power Out			1281.8	
Thermal Power				
Condensate Heat Rejection		1183.5		
Stack Loss		266.8		
Coal Drying		57.7		
Rejected in Solid Waste		21.3		
Internal Electric Power Not Regen.		48.7		
Other Losses to Ambient		144.4		
Thermal Power Out			1722.4	
Total Power Out				3004.2
Unaccounted For				9.7

- -

Efficiency

Excluding Seed Reprocessing	42.53%
Including Seed Reprocessing	42.33%

ENERGY FLOW SUMMARY

Case Number: 2.18

Changes from Reference Case: Regeneratively, Air-Cooled, Preheat Combustor; 1300 F Air & Flue Gas Into Preheat Combustor

---

Power In (MW)				
Raw Coal to Air Heater Subsystem			867.1	
Raw Coal to Main Combustor			1800.2	
Total Power In				2667,3
Power Out (MW)				
Electric Power				
MHD Generator		658.3		
Steam Turbines				
Shaft Power	765.3			
Main Compressor	-171.6			
Generator Loss	- 8.7			
Net Steam Power		585.0		
Plant Internal Power				
Inverters	9.9			
O_2 Plant	0.0			
Electrical Auxiliaries	52.8			
Internal Power		-62.7		
Electric Power Out			1180.6	
Thermal Power				
Condensate Heat Rejection		1031.8		
Stack Loss		218.7		
Coal Drying		63.3		
Rejected in Solid Waste		14.1		
Internal Electric Power Not Regen.		44.6		
Other Losses to Ambient		101.9		
Thermal Power Out			1474.4	
Total Power Out				2655.0
Unaccounted For				12.3

Efficiency

Excluding Seed Reprocessing	44.26%
Including Seed Reprocessing	44.03%

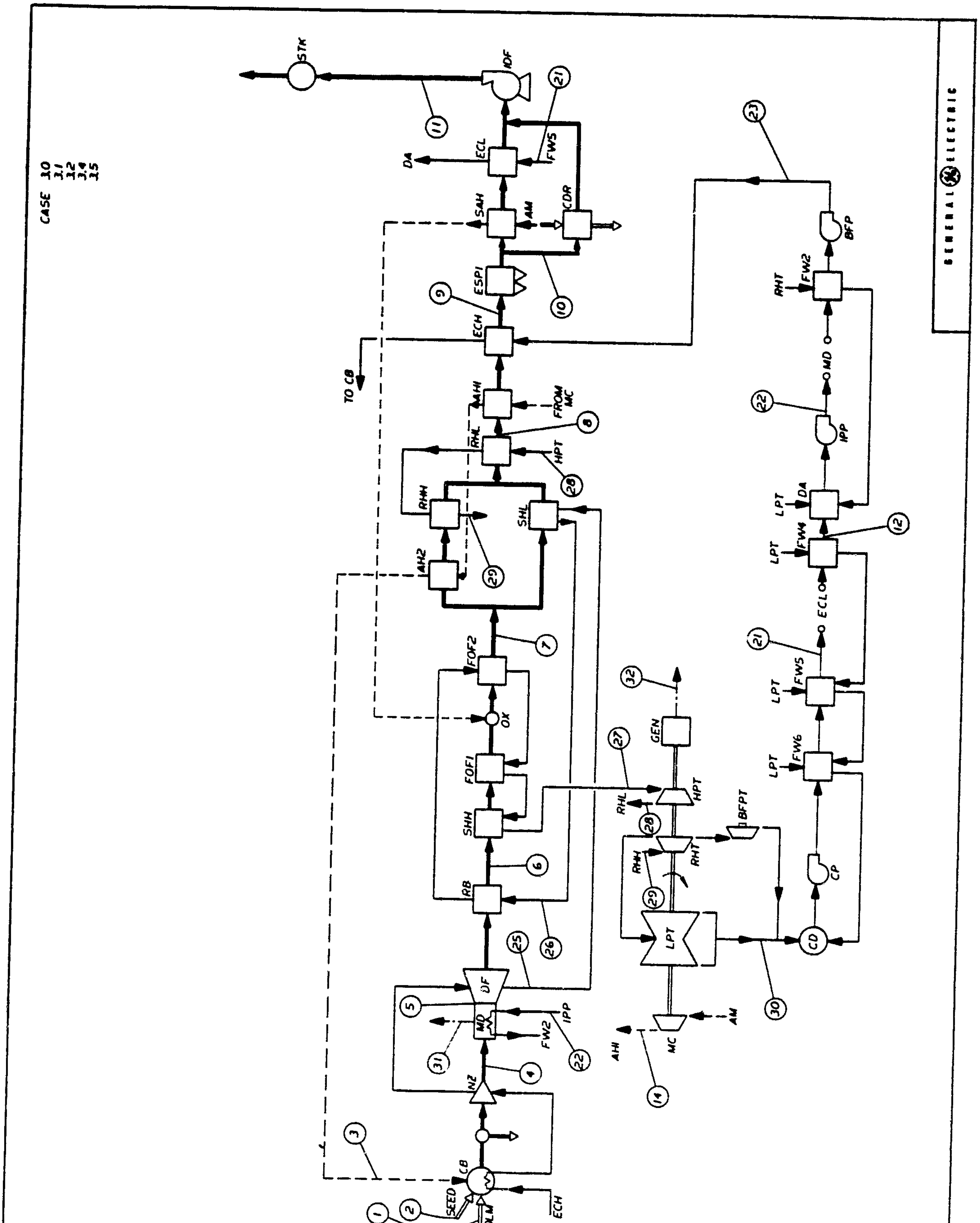
BASE CASE 3

INDEX

Case Number	Overall Efficiency	System Diagram	State Points	Energy Flow Summary
3.0	✓	✓	✓	✓
3.1	✓	✓		✓
3.2	✓	✓		✓
3.4	✓	✓		✓
3.5	✓	✓		✓

Overall Plant Efficiency
Base Case 3

Case No.	Parameter Variation	Efficiency, % Without Seed Reprocessing	Efficiency, % With Seed Reprocessing
3.0	Reference System 1-Stage Cyclone; 70% Slag Removal	43.25	42.91
3.1	Illinois #6 1-Stage Cyclone, 70% Slag Removal	43.72	42.66
3.2	2-Stage Cyclone 85% Slag Removal	43.23	42.89
3.4	1100 F Preheat	42.90	42.55
3.5	8 Tesla Channel	44.42	44.07


CASE 1.0
3.1
3.2
3.4
3.5
STK
IDF
11
21
ECL
FWS
DA
SAH
AM
CDR
ESP1
9
10
ECH
TO CB
AH1
FROM MC
RHL
HPT
8
28
RHH
SHL
29
AH2
FOF2
7
OX
FOF1
SHH
RB
6
26
25
DF
5
MD
IPP
FW2
22
31
NZ
4
3
CB
SEED
COLM
ECH
1
2
32
GEN
27
RHL
HPT
28
RHH
RHT
BFPT
29
LPT
30
MC
AM
AH1
14
CD
CP
LPT
FW6
FW5
21
ECL
FW4
12
DA
IPP
22
MD
RHT
FW2
BFP
23
GENERAL ELECTRIC

STATE POINTS, CASE 3.0

LOCATION	T (K)/(F)	PSIA	M (Kg/SEC)	E (MW)
1	220/373	-	-	-
2	220/373	-	-	-
3	978/1300	137.2	378.1	279.0
4	2816/4790	136.7	486.4	2608.0
5	2351/3737	17.7	486.4	1894.0
6	1866/2899	14.65	486.4	1488.6
7	1330/1935	14.54	609.7	897.1
8	692/725	14.36	609.7	400.8
9	521/478	14.22	609.7	282.
10	528/491	14.22	285.7	134.5
11	384/231	14.75	590.7*	191.2
14	608/634	150.4	378.1	124.4
21	361/190	157.	538.4	-
12	401/261	137.	538.4	-
22	423/301	400.	562.7	-
23	492/427	3860.	562.7	-
25	606/631	3670.	562.7	-
26	653/715	3610	562.7	-
27	811.1/1000	3500.	562.7	-
28	591/604	768.	552.5	-
29	811/1000	691.	557.3	-
30	314/106	2.3 "Hg	421.6	-
31	-	-	-	560.1
32	-	-	-	654.6

*Excludes Coal Drying Moisture

ENERGY FLOW SUMMARY

Case Number: 3.0

Changes from Reference Case: Reference Case

- -

Power In (MW)

Raw Coal to Air Heater Subsystem		0.0	
Raw Coal to Main Combustor		2518.4	
Total Power In			2518.4

- -

Power Out (MW)

Electric Power				
MHD Generator		560.1		
Steam Turbines				
Shaft Power	791.2			
Main Compressor	-125.1			
Generator Loss	11.5			
Net Steam Power		654.6		
Plant Internal Power				
Inverters	-8.4			
O_2 Plant	-84.9			
Electrical Auxiliaries	-32.1			
Internal Power		-125.4		
Electric Power Out			1089.3	
Thermal Power				
Condensate Heat Rejection		1037.8		
Stack Loss		154.7		
Coal Drying		54.2		
Rejected in Solid Waste		15.1		
Internal Electric Power Not Regen.		122.4		
Other Losses to Ambient		34.3		
Thermal Power Out			1418.5	
Total Power Out				2507.8
Unaccounted For				10.6

- -

Efficiency

Excluding Seed Reprocessing	43.25%
Including Seed Reprocessing	42.91%

ENERGY FLOW SUMMARY

Case Number: 3.1

Changes from Reference Case: Illinois #6 Coal

Power In (MW)

Raw Coal to Air Heater Subsystem			0	
Raw Coal to Main Combustor			2520.0	
Total Power In				2520.0

Power Out (MW)

Electric Power				
MHD Generator		613.5		
Steam Turbines				
Shaft Power	777.1			
Main Compressor	-142.5			
Generator Loss	- 11.2			
Net Steam Power		623.3		
Plant Internal Power				
Inverters	-9.2			
O_2 Plant	-84.7			
Electrical Auxiliaries	-32.1			
Internal Power		-126.0		
Electric Power Out			1101.8	
Thermal Power				
Condensate Heat Rejection		1019.2		
Stack Loss		160.0		
Coal Drying		45.5		
Rejected in Solid Waste		14.5		
Internal Electric Power Not Regen.		123.0		
Other Losses to Ambient		34.0		
Thermal Power Out			1396.7	
Total Power Out				2507.0
Unaccounted For				13.

Efficiency

Excluding Seed Reprocessing	44.08%
Including Seed Reprocessing	43.02%

ENERGY FLOW SUMMARY

Case Number: 3.2

Changes from Reference Case: 2nd Stage Cyclone Combustor

Power In (MW)				
Raw Coal to Air Heater Subsystem			0	
Raw Coal to Main Combustor			2518.0	
Total Power In				2518.0
Power Out (MW)				
Electric Power				
MHD Generator		585.1		
Steam Turbines				
Shaft Power	779.4			
Main Compressor	-138.8			
Generator Loss	- 11.1			
Net Steam Power		629.5		
Plant Internal Power				
Inverters	-8.8			
O_2 Plant	-84.9			
Electrical Auxiliaries	-32.1			
Internal Power		-125.8		
Electric Power Out			1088.8	
Thermal Power				
Condensate Heat Rejection		1022.3		
Stack Loss		169.5		
Coal Drying		54.2		
Rejected in Solid Waste		15.3		
Internal Electric Power Not Regen.		122.8		
Other Losses to Ambient		34.2		
Thermal Power Out			1418.3	
Total Power Out				2507.1
Unaccounted For				11.3

Efficiency

Excluding Seed Reprocessing	43.23%
Including Seed Reprocessing	42.89%

ENERGY FLOW SUMMARY

Case Number: 3.4

Changes from Reference Case: Air Preheat Reduced from 1300 F to 1100 F

Power In (MW)				
Raw Coal to Air Heater Subsystem			0	
Raw Coal to Main Combustor			2561.5	
Total Power In				
Power Out (MW)				
Electric Power				
MHD Generator		550.6		
Steam Turbines				
Shaft Power	813.6			
Main Compressor	-126.8			
Generator Loss	- 11.9			
Net Steam Power		674.9		
Plant Internal Power				
Inverters	-8.3			
O_2 Plant	-86.3			
Electrical Auxiliaries	-32.1			
Internal Power		-126.7		
Electric Power Out			1098.8	
Thermal Power				
Condensate Heat Rejection		1067.7		
Stack Loss		151.3		
Coal Drying		55.4		
Rejected in Solid Waste		15.2		
Internal Electric Power Not Regen.		122.3		
Other Losses to Ambient		34.5		
Thermal Power Out			1446.4	
Total Power Out				2545.2
Unaccounted For				16.3
Efficiency				
Excluding Seed Reprocessing	42.90%			
Including Seed Reprocessing	42.55%			

ENERGY FLOW SUMMARY

Case Number: 3.5

Changes from Reference Case: 8 Tesla Magnet

- -

Power In (MW)

Raw Coal to Air Heater Subsystem			0	
Raw Coal to Main Combustor			2518.4	
Total Power In				2518.4

- -

Power Out (MW)

Electric Power				
MHD Generator		658.3		
Steam Turbines				
Shaft Power	749.2			
Main Compressor	-151.6			
Generator Loss	- 10.3			
Net Steam Power		587.3		
Plant Internal Power				
Inverters	-9.9			
O_2 Plant	-84.9			
Electrical Auxiliaries	-32.1			
Internal Power		-126.9		
Electric Power Out			1118.7	
Thermal Power				
Condensate Heat Rejection		983.5		
Stack Loss		179.5		
Coal Drying		54.2		
Rejected in Solid Waste		15.1		
Internal Electric Power Not Regen.		123.8		
Other Losses to Ambient		32.8		
Thermal Power Out			1388.9	
Total Power Out				2507.6
Unaccounted For				10.8

- -

Efficiency

Excluding Seed Reprocessing	44.42%
Including Seed Reprocessing	44.07%

APPENDIX C

CALCULATION OF STATE AND TRANSPORT PROPERTIES OF PRODUCTS OF COAL COMBUSTION

APPENDIX C

CALCULATION OF STATE AND TRANSPORT PROPERTIES OF PRODUCTS OF COAL COMBUSTION

The Coal Combustion Equilibrium (CCE) Computer Code is a general purpose chemical equilibrium program which is applied to coal combustion products[1]. The calculation determines the composition and thermodynamic properties of a system consisting of a gaseous phase in equilibrium with any number of species which are ideally mixed in the sense that they obey Raoults law. A Margules-type model can be used to describe the ternary system of potassia, alumina and silica.

The primary data base for the program is the computer tape version of the JANAF thermochemical tables. This has been augmented using data from other sources and also by revisions based on improved values of the heat of formation for certain species. The updated thermodynamic data are given in Reference 1.

Transport properties (viscosity and thermal conductivity) of the heavy particles are computed using the method of Hirshfelder, Curtis, and Bird[2] and assuming that a Lennard-Jones 6-12 potential applies for interactions between colliding species. The individual species are combined using Wilke's rule[3] to obtain properties for the combustion product mixture.

The model of Demetriades and Argyropoulos[4] for electron transport properties has also been programmed as a subroutine for use with Code CCE. Cross-section integrals which describe the interaction between electrons and neutral species were taken from Reference 3. Coulomb interactions were computed using the collision integrals for a screened coulomb potential taken from Mason, Munn and Smith[6].

A sample output page from a typical CCE calculation is shown in Table C-1. The temperature and pressure are listed at the top of the page followed by the equilibrium composition in moles per 100 Kg of coal. The gas phase species are listed first followed by three condensed phases, water soluble, graphite and glass (slag). The thermodynamic properties of each phase and mixture are given followed by the transport and electrical properties of the gas. All units are MKS.

<u>References for Appendix C</u>

1. Cook, C. S., et.al., "Evaluation of Technical Feasibility of Closed Cycle Non-Equilibrium MHD Power Generation with Direct Coal Firing Final Report,: GE Report to be Published.

2. Hirschfelder, J. O., Curtiss, C. F. and Bird, R. B., <u>Molecular Theory of Gases And Liquids</u>, Wiley, New York, 1954.

3. Bird, R. B., Stewart, W. E. and Lightfoot, E. N., <u>Transport Phenomena,</u> Wiley, New York, 1960.

4. Demetriades, S. T. and Argyropoulos, G. S., Phys. of Fluids, <u>9</u>, 2136, 1966.

5. Spencer, F. E. and Phelps, A. V., "Momentum Transfer Cross-Sections and Conductivity Integrals for Gases of MHD Interest," 15th Symposium EAMHD, Philadelphia, May 24-25, 1976.

6. Mason, E. A., Munn, R. J. and Smith, F. J., Phys. of Fluids, <u>10</u>, 1827, 1967.

Table C-1. Sample CCE Output

```
   TEMPERATURE  2700.0          PRESSURE    3.9964

 EQUILIBRIUM COMPOSITION.
      3.218097E 02  0.            0.            1.814474E-04
ALO2    9.136661E-04  CO2     4.052483E 01  CA      8.110235E-04  CL2     2.465406E-20  FE      3.363782E-03  H2      1.535789E 00
K       1.182781E 00  N2      2.222640E 02  NA      8.459215E-03  O2      4.344078E 00  P2      4.736039E-14  S2      1.891509E-07
SIO2    1.377717E-02  A       2.662914E 00  E-      8.263731E-03
AL      3.495403E-06  ALO     1.754543E-04  ALO2H2  3.784820E-02  CO      1.668506E 01  CSO     5.077601E-06  CAO     7.884083E-04
CAOH    2.221380E-02  CL      7.547731E-09  FEO     3.231209E-03  FEO2H2  1.170975E-03  H       4.285958E-01  HCL     2.664163E-08
H2O     2.426498E 01  SH2     1.131732E-05  KCL     2.439115E-07  KO      3.278184E-02  KOH     1.190117E 00  N       2.167052E-04
NO      2.541766E 00  NO2     8.515437E-04  NACL    6.316986E-10  NAO     2.205921E-04  NAOH    3.483757E-03  O       6.705127E-01
OH      2.528639E 00  P       3.946212E-08  PH      2.104935E-09  PH2     1.279020E-10  PO      1.646601E-04  PO2     5.207451E-04
PS      3.932694E-10  S       3.804450E-04  SO      2.408882E-02  SO2     3.450541E-01  SO3     1.077562E-04  SI      1.333822E-08
SIO     6.232004E-02  AL+     1.146547E-11  ALO-    8.004538E-05  ALO2-   2.126069E-03  CN-     3.769565E-08  CO2-    1.728448E-05
CL-     3.331045E-09  K+      1.130175E-02  KO-     2.766140E-05  NO2-    7.853217E-06  NA+     2.619337E-06  O-      7.523554E-05
OH-     6.848709E-04  O2-     2.157515E-05
KCL     0.            KOH     0.            K2CO3   0.            K2SO4   0.            NACL    0.            NAOH    0.
NA2CO3  0.            NA2SO4  0.
C       0.
CAO     4.286176E-05  FEO     2.342523E-06  NAO.5   8.776449E-10  KO.5    5.823116E-08  SIO2    1.116564E-05  ALO1.5  1.250183E-04

 SPECIES             MOLES FREE ENERGY    ENTROPY ENTHALPY  HEAT OF FORM  WMOL        MASS
                                                JOULES                                KILOGRAMS

GAS               3.2181E 02 -2.2393E 07  8.7011E 04  5.6938E 06 -2.3365E 07  2.9353E 01  9.4459E 00
WATER SOL.        0.          0.          0.          0.          0.          0.          0.
GRAPHITE          0.          0.          0.          0.          0.          0.          0.
GLASS             1.8145E-04 -7.1223E 01  3.4676E-02 -1.0383E 02 -1.3498E 02  5.3013E 01  9.6191E-06
                  3.2181E 02 -2.2393E 07  8.7011E 04  5.6937E 06 -2.3365E 07  2.9353E 01  9.4459E 00

 GAS DENSITY  5.2948E-01 KG/M**3
 GAS VOLUME   1.7840E 01 M**3

    TRANSPORT PROPERTIES     ** M.K.S. UNITS **
  VISCOSITY  7.7660E-05 THERMAL CONDUCTIVITY  8.6811E-02

 ELECTRICAL PROPERTIES. DEBYE OPTION  0
SIGMA                9.646E 00 BETA1            -2.237E-02 BETA2            0.          THETA1           7.440E-06
LAMDA                3.604E-04 DEBYE LENGTH      2.306E-07 COLLISION FREQ.  6.135E 11
                         MAGNETIC INDUCTION
 B                   0.          2.0         4.0         6.0         8.0         10.0        12.0
CHI                  2.244E-02   2.243E-02   2.242E-02   2.241E-02   2.240E-02   2.239E-02   2.239E-02
HALL PARAMETER       0.          5.734E-01   1.147E 00   1.720E 00   2.294E 00   2.867E 00   3.440E 00
PSI                  1.071E-05   9.710E-06   7.579E-06   5.550E-06   4.036E-06   2.989E-06   2.269E-06
THETA2              -1.196E-06  -1.084E-06  -8.461E-07  -6.196E-07  -4.506E-07  -3.337E-07  -2.533E-07
THETA3               1.923E-07   1.743E-07   1.360E-07   9.961E-08   7.245E-08   5.364E-08   4.072E-08
ALAM2               -5.794E-05  -5.251E-05  -4.099E-05  -3.001E-05  -2.183E-05  -1.616E-05  -1.227E-05
ALAM3                9.315E-06   8.442E-06   6.590E-06   4.825E-06   3.509E-06   2.598E-06   1.973E-06
```

APPENDIX D

REPROCESSING OF SPENT SEED PRODUCED BY AN MHD/STEAM POWER GENERATING SYSTEM

APPENDIX D

REPROCESSING OF SPENT SEED PRODUCED BY AN
MHD/STEAM POWER GENERATING SYSTEM

This survey was prepared to introduce Hooker Chemical Company to seed reprocessing requirements as a supplement to an oral briefing.

I. Introduction

Rather stringent requirements for seed recovery and regeneration in coal-fired, open-cycle MHD/steam power generating systems are imposed by economic considerations. Briefly, it is generally acknowledged that the economic viability of any open-cycle power generating plant depends heavily on the efficient recovery of at least 95% of the seed added to the system. Based on the costs of alkali metal salts, potassium compounds such as K_2CO_3 appear to be prime candidates for use as seedant materials. However, it is possible that under certain circumstances, the economic advantages gained through the use of cesium compounds (lower combustion temperature) can more than offset the higher costs of these materials.

The present document reviews system requirements for reprocessing spent seed with emphasis placed on the following areas:

1. The chemical state of the spent seed as it is recovered from the MHD power generating system.
2. The disposal of sulfur from the reprocessing system.
3. The requirements of the reprocessed seed.

The work centers on potassium compounds but because of chemical similarities, the results and conclusions also apply to cesium.

This document represents the first step in evaluating seed reprocessing systems for PSPEC. The intent is to provide a basis for technical discussions between NASA LeRC, Hooker Chemical Company and GE. These discussions will lead to selection of the most promising seed reprocessing system which will then be evaluated on an economic basis.

II. Characteristics of Spent Seed

Seed which is injected into the combustor reaches thermodynamic equilibrium with the combustion gases and moves with the fluid through the entire flow train. To illustrate the seed behavior as a function of temperature, a simplified diagram of the interaction of potassium seed with combustion products of natural gas in the U-25 facility is shown in Figure D-1 (Reference 1). At high temperature, above 2500 K, the major seed species is gaseous, elemental potassium. As the flow cools, potassium reacts with water to form KOH vapor. Below about 1500 K, KOH reacts with CO_2 to form K_2CO_3. As the flue gases become saturated with these vapors, condensation occurs which results in an aerosol. At sufficiently low temperature, the liquid aerosol solidifies. Below about 440 K, carbonation of K_2CO_3 to form $KHCO_3$ takes place, and finally, at low temperature the $KHCO_3$ absorbs water.

From the above discussion, it becomes evident that the spent seed which is recovered from the power generating system will be of mixed composition. Further, the chemical composition will depend on the recovery point in the flow train. This is illustrated by seed recovery experience obtained in the low sulfur, U-25 system (Reference 1).

In the high temperature zone, where KOH is the major seed species, condensation from the gas phase occurs on the cold heat transfer surfaces. The condensed KOH then reacts with CO_2 from the combustion gases to form a hard, crust like deposit with a melting temperature of about 1000 K. The thermal and mass transport from the hot gases to the cold walls is such that the deposited layer becomes stablized in thickness and further condensate runs off as KOH liquid. This behavior allows the recovery of the spent seed

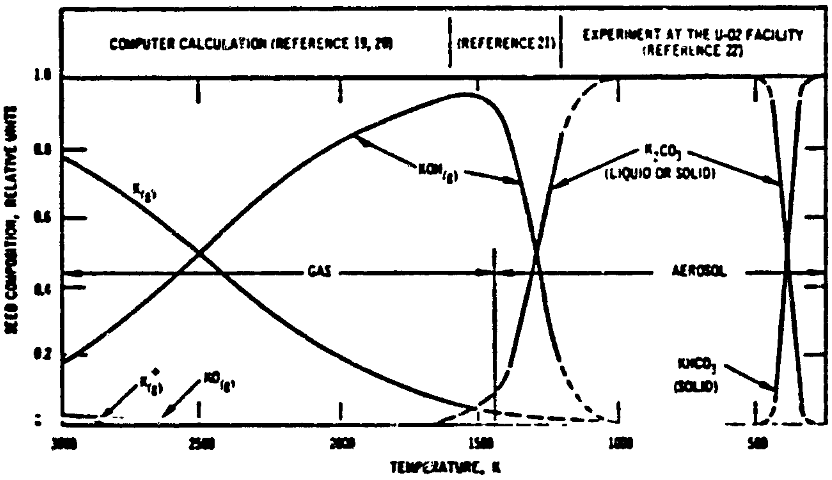


Figure D-1. Chemical Composition of Seed Species in the Sulfur Free U-25 Facility. Taken from Reference 1.

which runs to the bottom of the radiant and semiradiant sections of the flow train. About 40% of the total seed in the flow is recovered in this manner.

At lower temperatures where the seed occurs as a K_2CO_3 aerosol, interaction between the flow and convective tube banks causes K_2CO_3 deposition. These deposits must be removed mechanically or by water washing in order to prevent plugging of the flow passages. The spent seed which is recovered in this section consists of almost pure K_2CO_3 in either solid or aqueous form depending on the removal technique.

The remainder of the seed in the flow is removed at low temperature using wet or dry electrostatic precipitators, or wet venturi scrubbers. Because of the carbonation of K_2CO_3 by the CO_2 present in the flue gases, the temperature and method of final removal determines the composition of the recovered spent seed. Above about 440 K, the seed is recovered as K_2CO_3. However, as the temperature of recovery falls, the amount of $KHCO_3$ in the spent seed increases. Because $KHCO_3$ is formed at temperatures above the boiling point of seed/water solutions, wet methods will produce large amounts of $KHCO_3$ particularly at low solution concentrations, Dry electrostatic precipitation above 440 K results in almost pure K_2CO_3 as a product.

The major effect of sulfur from the fuel on seed phenomenology is the formation of K_2SO_4 at a temperature of about 1650 K. This behavior is illustrated in Figure D-2. Thus, in the radiant section of the furnace, one would expect K_2SO_4 deposits to accumulate on the walls until the

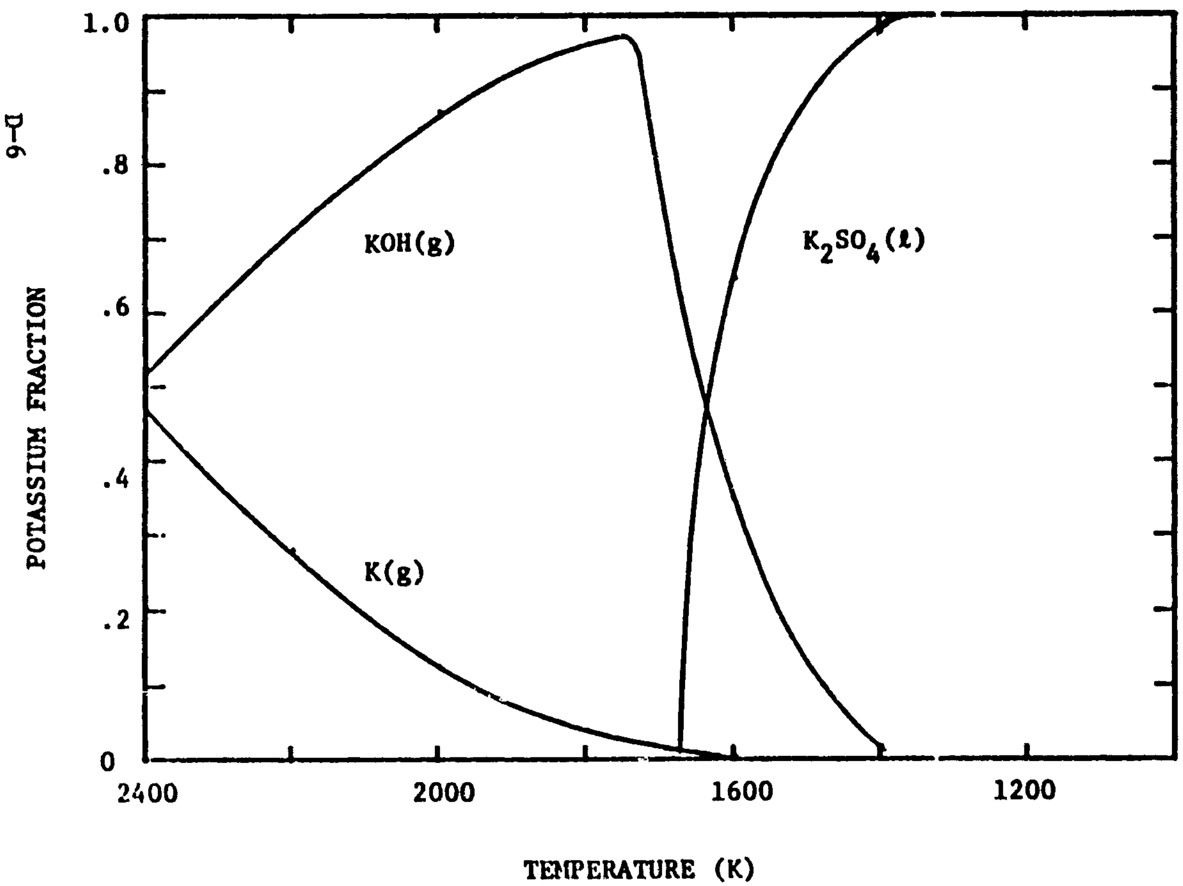


Figure D-2. Chemical Composition of Seed Species in Equilibrium with Combustion Products from a Sulfur Containing Coal

thickness of the deposits is stablized by heat and mass transfer effects. At this point further condensate will run off as a liquid. The difference between sulfur-free and sulfur-containing flow is the presence of K_2SO_4 in the liquid collected at the bottom of the flow train.

At lower temperature in the convective and particulate removal sections, K_2SO_4 will be collected in a manner similar to that described for K_2CO_3. For seed loadings in excess of that required for total sulfur removal, the additional seed will be collected as K_2CO_3.

Additional potassium sulfur compounds can be formed in the flow train under specific conditions. When the combustion products are fuel rich, K_2S rather than K_2SO_4 may be the preferred sulfur/potassium compound. Uncertainties in the thermodynamic properties of K_2S prevent quantification of this behavior. However, it is highly probable that the liquid seed effluent from the radiating section of the steam generating plant prior to injection of secondary air, will contain some K_2S. In addition, at low temperatures K_2SO_4 may continue to react with remaining sulfur oxides to form $K_2S_2O_7$ and $KHSO_4$. These compounds are expected to be present in the spent seed collected by the particulate removal equipment.

The presence of slag in the combustion gases has several major effects on the phenomenology of seed recovery because the condensation temperature of slag (~ 2100 K) is higher than that of potassium compounds. The cool surfaces in the radiative section of the steam generating

plant will be coated with slag deposits. As discussed previously, the deposits will be stabilized by heat and mass transfer effects such that further condensate will run off as a liquid. The temperature of the liquid layer is such that slag rather than spent seed is collected at the bottom of the slagging furnace. However, potassium is soluble in liquid slag and, if dissolved, cannot be recovered easily. Thus, potassium losses to liquid slag layers are expected.

At lower gas temperatures the slag will condense to flyash which travels through the flow train. These slag particles can act as nucleation centers for seed condensation. Thus, a mixture of K_2SO_4, K_2CO_3 and flyash is expected to be the major product recovered from slagging systems. Proceeding comments on the presence of other potassium compounds also apply to the spent seed/flyash system.

Table D-1 presents a summary of the composition of the spent seed which is recovered from different components of the downstream flow train as a function of combustion gas purity. In clean systems, KOH and possibly $KHCO_3$ are the major side products of K_2CO_3 interaction with the combustion gases. When the flue gases contain sulfur, this species reacts with the seed to produce a mixture of K_2SO_4 and K_2S in the fuel-rich sections of the flow train. After secondary combustion, the K_2S is converted to K_2SO_4. The deposits on the convective tube banks will consist of both K_2SO_4 and K_2CO_3 with the mixture ratio determined by the sulfur content of the fuel. At low temperature, $K_2S_2O_7$ and possibly $KHCO_3$ are formed and collected

as particulates. The presence of slag in the flow eliminates the collection of a spent seed liquid fraction in the high temperature radiant furnace. Rather, the collected liquid is slag which contains dissolved K_2O. In the downstream components, the spent seed will be collected as a mixture which contains flyash.

It is concluded from the above discussion that the reprocessing of spent seed will be connected with the specific design of the MHD/steam power generating system and that seed regeneration requirements may have a significant systems impact.

Table D-1. COMPOSITION OF SPENT SEED RECOVERED FROM DIFFERENT SECTIONS OF THE FLOW TRAIN AS A FUNCTION OF COMBUSTION GAS PURITY

FLOW TRAIN COMPONENT	COMBUSTION GAS IMPURITIES		
	CLEAN	SULFUR	SLAG AND SULFUR
TEMPERATURE OF LIQUID LAYER IN RADIATIVE SECTION	1000 K	1500 - 1650 K	~ 2100 K
COMPOSITION OF LIQUID REMOVED FROM RADIATIVE SECTION	KOH	K_2SO_4, K_2S	SLAG AND DISSOLVED K_2O
COMPOSITION OF DEPOSITS REMOVED FROM TUBE BANKS IN CONVECTIVE SECTION	K_2CO_3	K_2SO_4, K_2CO_3	K_2SO_4, K_2CO_3, FLYASH
COMPOSITION OF PARTICULATES COLLECTED USING A DRY ESP	K_2CO_3	K_2SO_4, K_2CO_3, $K_2S_2O_7$	K_2SO_4, K_2CO_3, $K_2S_2O_7$, FLYASH
COMPOSITION OF PARTICULATES COLLECTED USING WET METHODS	$KHCO_3$	K_2SO_4, $KHCO_3$, $K_2S_2O_7$	K_2SO_4, $KHCO_3$, $K_2S_2O_7$, FLYASH

III. Requirements for Seed Reprocessing

Ninety-five percent of the spent seed must be recovered and reused in any viable open cycle MHD/steam power generating system. This requirement places severe restrictions on seed reprocessing facilities. In the case of sulfur-containing fuels, a major portion of the recovered spent seed is in the form of potassium/sulfur compounds. Reprocessing of these compounds to remove sulfur is required and must be efficient as to energy consumption and product yield.

The products of seed reprocessing are not fixed but will be determined by economic considerations. For example, the sulfur which must be removed from the spent seed can either be reduced to elemental sulfur, recovered as sulfuric acid or discarded as an insoluable sulfate. The first two options provide marketable products whereas, the disposal of sulfates may cause problems. Considering these three possible options in order, elemental sulfur is easy to store and ship. If a market for the sulfur cannot be found locally, and the shipping costs are prohibitive, the material can be used as a clean land fill. The problem with elemental sulfur as a final product is economic in nature. Additional energy and equipment are required for production of this material in contrast to throw-away systems. Sulfuric acid is also cheaper to produce than elemental sulfur. This material is an ideal product if a local market can be found. However, storage and shipment of sulfuric acid cause difficulties and disposal of unwanted sulfuric acid is difficult.

The throw-away option is appealing at the plant site because it is efficient in the use of energy and equipment. However, the product (usually $CaSO_4$) is not marketable, is hard to handle and has caused problems in disposal. In addition, a cheap source of calcium is required in that it is discarded with the sulfur in the coal on a 1:1 molal basis. From the above discussion, it can be seen that the most desirable seed reprocessing system is sensitive to plant location, local markets and disposal facilities.

In parallel to the question of the best final product for sulfur, the form of the recovered seed need not be K_2CO_3. The only actual requirement is that the seed be handable, combustible and not produce polluting species. Thus, K_2CO_3, KOH, K_2O and organic potassium salts are all acceptable as feedstock for the combustor. Some compounds might have slight thermal or material handling advantages but these are probably secondary if a major benefit to the seed reprocessing facility is realized by producing one compound in contrast to another.

An additional requirement for seed reprocessing is removal of trace impurities such as sodium, iron, other metals and chlorides which are entrapped by the spent seed as it passes through the steam generating plant. Some of these impurities are present in coal while others are produced by the interaction of the seed with the construction materials of the steam plant. If the impurities are not removed, they may accumulate in the seed charge over a period of time and produce an undesirable loss in plant efficiency.

It is desirable to decouple the seed reprocessing facility from the main flow train because of outage problems. It is not acceptable for the power generating system to rely on the operational status of the seed reprocessing facility. Consequently, provisions to store spent seed as well as reprocessed seed must be made. Because of the differences in the composition of spent seed which is recovered from various sections of the flow train, several storage areas may be required depending upon the seed reprocessing system which is selected.

To summarize the discussion on seed reprocessing requirements, the selected facility must be energy efficient with a high product yield. The final form of the sulfur from the coal need not be a specific chemical compound but an option to produce elemental sulfur, sulfuric acid or a discardable sulfur compound exists. The selection of this option is site specific. In a like manner, the chemical form of the reprocessed seed is not fixed and will be selected by economic considerations. A major requirement is seed purity to prevent build up of unwanted elements in the recycled seed flow. Lastly, it is desirable to decouple the operation of the steam power generating system from seed reprocessing.

IV. Summary and Conclusions

A review of systems requirements for reprocessing spent seed has been undertaken. Emphasis was placed on the chemical state of the spent seed as it is recovered from the MHD/steam power generating system. Reprocessing requirements were also examined in order to define desirable sulfur and products and the chemical characteristics of the reprocessed seed.

An examination of the spent seed which will be recovered in various components of the steam generating flow train led to the conclusion that seed reprocessing is directly connected to the specific design of the power plant. Two major differences between slagging and non-slagging systems are identified. First, spent seed collected from slag containing systems will be a mixture of flyash, K_2SO_4 and K_2CO_3. $KHCO_3$ will be formed from K_2CO_3 if the seed/flyash particles are removed at temperatures below 440 K. Second, spent seed obtained from slag-free systems will be collected in two fractions: a) liquid K_2SO_4 mixed with K_2S will be collected in the radiant boiler and b) a mixture of K_2SO_4 and K_2CO_3 ($KHCO_3$ for low particulate removal temperatures) will be collected in the convective section and particle removal equipment. These differences indicate that seed reprocessing facilities may be different for the slagging and non-slagging cases.

Requirements for seed reprocessing systems are quite flexible. The equipment must be efficient with a high product yield. However, the final form of the sulfur processed from the spent seed is optional and will be

defined on a site specific economic basis. The chemical composition of the processed seed is also optional with the major requirements being an easily handlable material which is combustible. Trace elements cannot be allowed to accumulate in the processed seed. Finally, the operation of the power generating system must be independent of the seed reprocessing facility.

References-Appendix D

1. Petrick, M. and Shumyatsky, B. Ya., Editors, "Open Cycle Magnetohydrodynamic Electrical Power Generation," DOE TR-119, Argonne National Laboratory, 1978, Chapter 12.

APPENDIX E

MHD SEED REGENERATION PROCESS EVALUATION

APPENDIX E

MHD SEED REGENERATION PROCESS EVALUATION

This appendix is a report to GE from the Hooker Chemical Company.

For a discussion of how the data was applied to PSPEC cases, see Section 3.9.

hooker

February 28, 1979

MHD SEED REGENERATION PROCESS EVALUATION

for Electrochemical and Formate processes.

by George T. Miller

In response to a request by Dr. Fred N. Alyea of General Electric Co., a preliminary cost comparison was made for the electrochemical process and the formate process for recovering potassium values from MHD seed materials. The problem centers about the recovery of better than 95% of the potassium values separate from the sulfur values which are introduced with the coal.

A review of reports supplied by Dr. Alyea shows a wide range of plant sizes used for the economics of seed recovery and thus fail to provide for direct comparison of the various seed regeneration processes and for the economic effect of fuels with various sulfur content. The MHD plant of this request, based on 1135 MWe at 43.2% efficiency (2,627 MWt) using Montana Rosebud coal at 1.1% sulfur, was not noted in other reports provided. Extension of scope to Illinois #6 coal containing 3.55% sulfur was also requested. Not requested, but germaine, would be processing the minimum sulfur values while venting the maximum SO_2 per EPA emission limits- an emergency provision.

Therefore to effectively compare processes, the following plant capacities for feed recycle are compared:-

code: M = Montana Rosebud coal (1.1% S)
I = Illinois # 6 coal (3.55% S)
N = no SO_2 vented in stack
V = Vent max. SO_2 per EPA emission limits (1.2# $SO_2/10^6$BTU)
2.6 = 2,627 MWt MHD plant
2 = 2,000 MWt MHD plant
1 = 1,000 MWt MHD plant

Plant Code	#/hr K_2SO_4 processed	Plant Code	#/hr K_2SO_4 processed
IN-2.6	136,922 (requested)	MN-2.6	46,697 (requested)
IV-2.6	107,282	MV-2.6	17,057
IN-2	105,325	MN-2	35,921
IV-2	82,525	MV-2	13,121
IN-1	52,662	MN-1	17,960
IV-1	41,262	MV-1	6,560

Assuming the seed regeneration plant can handle approximately 10% more than theory, data for plant capacities from 7,500 to 150,000 # K_2SO_4/hour are required. Computations at 7,500, 20,000, 50,000, 100,000 and 150,000 # K_2SO_4/hour will provide a plot from which comparisons can be made.

The basis for computations per Dr. Alyea's request, with addenda to include some stack venting, is as follows:

1135 MWe MHD plant @ 43.2% efficiency = 2627 MWt

If Montana Rosebud Coal is used @ 1.1 % S = 8,588 #S/hour in 780,747 #coal/hour.

For condition MN-2.6-- no S is vented and 8,588#S/hour requires processing 46,697 # K_2SO_4/hour.

For condition MV-2.6 in which the max. permissible SO_2 is vented (5,451# S/hour), 17,057# K_2SO_4 per hour must be processed.

If Illinois #6 coal is used @ 3.55 % S = 25,181 # S/hour in 709,371 # coal/ hour.

For condition IN-2.6 --- no sulfur is vented and 25,181 # S/ hour requires processing 136,922 # K_2SO_4/ hour.

For condition IV-2.6 in which the maximum permissible SO_2 is vented (5,451 # S/ hour), 107,282 # K_2SO_4 per hour must be processed.

SUMMARY OF RESULTS:

For Pounds per Hour of K_2SO_4 Regenerated of:-

	7,500	20,000	50,000	100,000	150,000
Electrochemical Process					
Total Capital (Millions)	5.7	12.5	24	45	63
Net energy debit (MWe)	4.4	11.6	29.1	58.2	88.1
@ 20mil/KWH = $/hr.	88	232	582	1164	1762
Material debit					
$Ca(OH)_2$ debit	16	44	108	217	327
O_2+H_2+Gypsum Credit					
** Total Cost (debit) of above in $/hour	104	276	690	1381	2089
Formate Process					
Total Capital (Millions)	3.3	7.5	15.3	28	40.5
Net energy debit (MWe)	0.37	0.75	1.19	2.39	3.58
@ 20 mil/KWH = $/hr.	7.	15.	24.	48.	72.
Material debit					
$O_2+Ca(OH)_2$ debit	69.	182.	452.	914.	1368.
** Total Cost (debit) of above in $/hour.	76.	197.	476.	962.	1440.

**Note: the total cost includes materials and energy, not capital.

CONCLUSIONS:

The results presented are preliminary and are to function as an initial guide only. The formate reactors, using a slurry, may require many more units than those computed from experience based on NaOH, although what is believed to be a reasonable factor was used. The electrochemical cells must be modified for this application and again what is believed to be a reasonable factor was used. Reduction to practice is required to clarify these values.

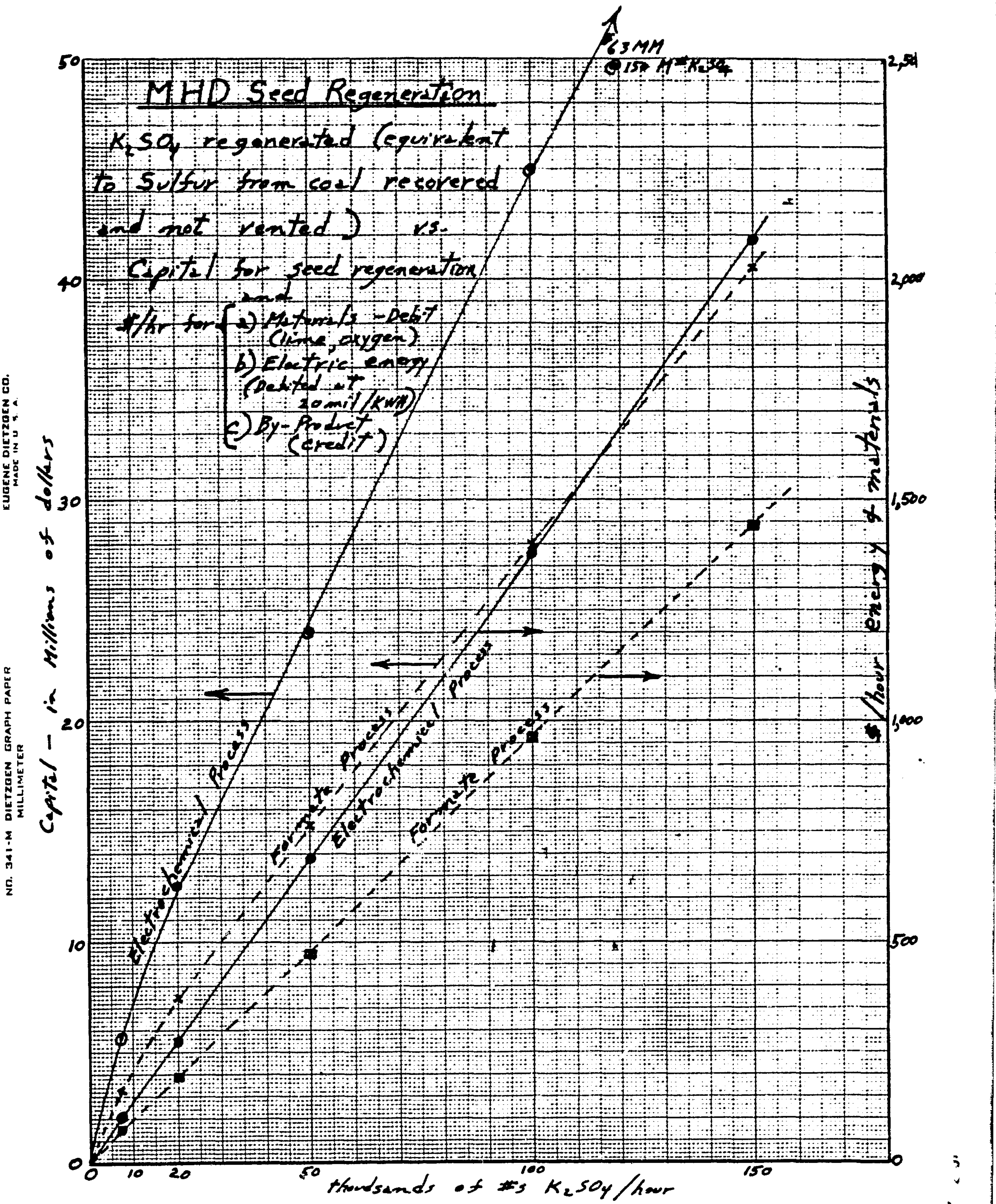
It is obvious that the computations would suggest that the formate process is preferred for both capital and operating material+ energy costs. It is pointed out that no value or disposal costs have been

ascribed to the $CaSO_4$ plus ash filter cake. This can be dumped or sand added and the mix reacted in a kilm to produce cement clinker and recover sulfuric acid.

The attached plot permits a cost analysis versus sulfur (as K_2SO_4) for an extensive range of plant sizes. To convert from pounds sulfur per hour to the pounds potassium sulfate per hour, for purposes of comparing other data, multiply by 5.44.

The attached addenda gives details of the electrochemical process and the formate process that were used in arriving at the data in this report.

George T. Miller
Scientist


MHD Seed Regeneration
K_2SO_4 regenerated (equivalent to Sulfur from coal recovered and not vented) vs.
Capital for seed regeneration and
\$/hr for a) Materials - Debit (lime, oxygen) b) Electric energy (Debited at 20 mil/KWH) c) By-Product (credit)
63 MM
@ 150 M# K_2SO_4
Electrochemical Process
Formate process
Electrochemical Process
Formate process
Capital - in Millions of dollars
\$/hour energy & materials
thousands of #s K_2SO_4/hour
0
10
20
30
40
50
0
10
20
50
100
150
500
1,000
1,500
2,000
2,500
EUGENE DIETZGEN CO. MADE IN U.S.A.
NO. 341-M DIETZGEN GRAPH PAPER MILLIMETER

ELECTROCHEMICAL PROCESS

for seed regeneration from MHD

The overall electrochemical reaction is:-

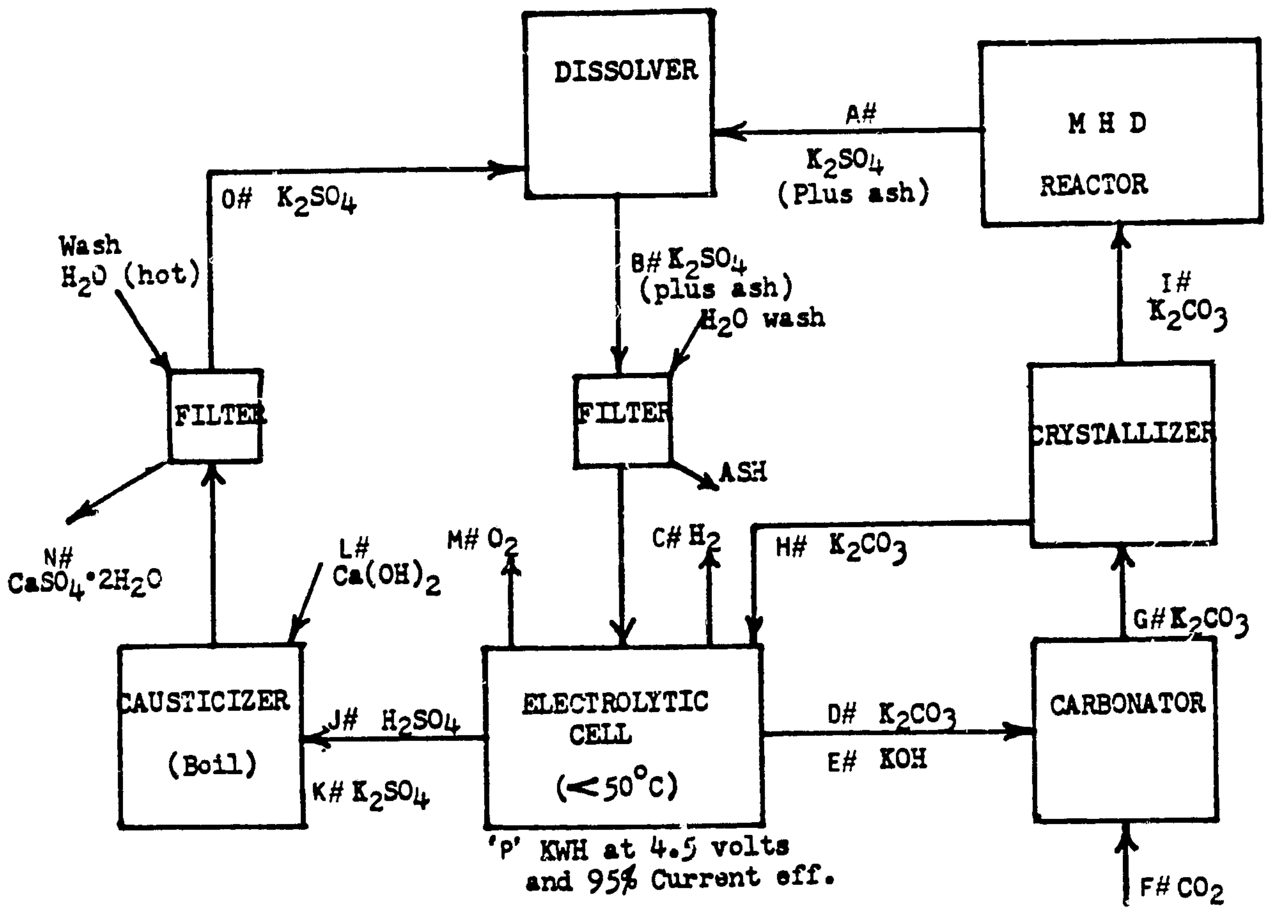
$K_2SO_4 + 3\ H_2O \longrightarrow 2\ KOH + H_2SO_4 + H_2 + 1/2\ O_2$

By converting the KOH into the carbonate, the potassium values can be separated by crystallization instead of requiring extensive evaporation. The mother liquor is then recycled to the catholyte. The hydrogen from the cathode compartment could be fed to a fuel cell and recover, as electricity, approximately 70% of the thermal energy or sold as a chemical at a maximum profit if a market can be found or simply used as a fuel and recover approximately 43 % of the thermal energy via the MHD route. For this preliminary cost analysis, the hydrogen was recycled as a fuel at a value of 30.5 MWt per ton of hydrogen. The oxygen was also recycled to the MHD reactor and given a value of 0.2 MWt per ton of oxygen as it enriched the combustion atmosphere and achieved higher combustion temperatures. The sulfuric acid plus potassium sulfate stream from the anolyte compartment was reacted with lime to produce pure calcium sulfate hydrate with a value of approximately \$7./ton and a recycling K_2SO_4 stream.

The electrolytic cells are modified H-2 and H-4 diaphragm cells. Cost for the cells, bus bars, piping, cranes, building, rectifiers, etc. was based on multiplying the Hooker costs by a factor to reflect the modifications. A voltage of 4.5 volts was used to compute power. The cells for the 7,500#/hr K_2SO_4 rate were sized at 80 KA, 16 in number. All other cells were 140 KA cells using 22 cells for 20,000 #/hr; 53 cells for 50,000 #/hr.;106 cells for 100,000 #/hr. and 160 cells for 150,000# K_2SO_4/hr.

All other costs of major equipment were computed from 1968 cost data as recorded in Chemical Engineering March 24, 1969 page 114 and adjusted to 1979 costs using a factor of 2.08.

SEED REGENERATION BY ELECTROLYSIS



Letter reference is to flow sheet on Page-2. All weights are expressed in pounds per hour.

A	K_2SO_4	7,500	20,000	50,000	100,000	150,000
B	K_2SO_4	18,750	50,000	125,000	250,000	375,000
C	H_2	86	230	575	1,151	1,727
D	K_2CO_3	5,953	15,874	39,685	79,369	119,054
E	KOH	4,821	12,856	32,140	64,280	96,419
F	CO_2	1,905	5,081	12,703	25,405	38,107
G	K_2CO_3	11,905	31,748	79,370	158,738	238,107
H	K_2CO_3	5,953	15,874	39,685	79,369	119,054
I	K_2CO_3	5,953	15,874	39,685	79,369	119,054
J	H_2SO_4	4,226	11,270	28,175	56,351	84,527
K	K_2SO_4	11,253	30,008	75,021	150,043	225,064
L	$Ca(OH)_2$	3,197	8,525	21,313	42,626	63,939
M	O_2	690	1,841	4,603	9,207	13,811
N	$CaSO_4 \cdot 2H_2O$	7,442	19,847	49,618	99,233	148,849
O	K_2SO_4	11,253	30,008	75,021	150,043	225,064
P	KWH/Hr	4,987	13,300	33,250	66,496	99,744
	K Amp/Hr	1,108	2,956	7,389	14,777	22,165

CREDITS & DEBITS:-	(per hour)				
O_2 @ .2MWt/ton MWt----	.07	.18	.46	.92	1.38
H_2 @ 30.5 MWt/ton MWt----	1.32	3.52	8.80	17.6	26.4
MWt credit for H_2 & O_2	1.39	3.70	9.26	18.52	27.78
MWe credit for H_2 & O_2	0.7	1.9	4.6	9.3	13.9
MWe debit for electrolysis	5.0	13.3	33.3	66.5	100.
MWe debit for agitators	.07	.22	.52	1.04	1.56
MWe debit-NET	4.37	11.6	29.1	58.2	88.1
(At 20mil/KWH- $ for MWe debit)	($87.)	($232.)	($582.)	($1164.)	($1762.)
$Ca(OH)_2$ debit @ $35/ton CaO	($43.)	($113.)	($282.)	($564.)	($848.)
Gypsum credit @ $7./ton	$26.	$69.	$174.	$347.	$521.

CAPITAL:-

Capital for individual items expressed as 'bare module' which includes all costs of equipment, installation, piping, steel, electrical, concrete, labor, site and similar direct and indirect costs but does not include contingency and contractor fees. The contingency and contractor fees will be added to the sum of the individual components to get the total module cost, the module being the seed regeneration plant. Note: 'M' = thousand and 'MM' = million in the following.

For #/hour K_2SO_4 regeneration rates of:- (Point A in flow sheet)

	7,500	20,000	50,000	100,000	150,000
Dissolvers:					
1 hr. retention	26 M gal.	70 M gal.	175 M gal	350 M gal	525 M gal
10% solution - 20%	100 H.P.	x3	x7	x14	x21
leeboard- steel	agitator				
	\$51,000	\$153,000	\$357,000	\$714,000	\$1,071,000
Filter:					
Rotating drum					
#ash/hr. removed	375	1,000	2,500	5,000	7,500
(5% of K_2SO_4)					
Ft^2 @ 13#/hr/ft^2	29	77	192	385	577
	\$39,000	\$72,000	\$128,000	\$199,000	\$256,000
Electrolytic :					
At 4.5 volts/	16 cells	22 cells	53 cells	106 cells	160 cells
cell- includes	@ 80KA	@ 140 KA	@ 140 KA	@ 140 KA	@ 140 KA
rectifiers	\$3.4 MM	\$7.2 MM	\$13 MM	\$25 MM	\$35 MM
Carbonators:					
$\#CO_2$/hour	2,000	5,500	13,000	27,000	40,000
Ft^3@4ft^3/$\#CO_2$/hr	8,000	22,000	52,000	108,000	160,000
Bubble plates @					
2 ft.height-					
	\$0.5 MM	\$1.4 MM	\$3.3 MM	\$6.6 MM	\$10. MM
Crystallizers:					
# K_2CO_3/hour	6,000	16,000	40,000	80,000	120,000
	\$53 M	\$90 M	\$150 M	\$219 M	\$274 M
Filter for K_2CO_3:					
# $2K_2CO_3 \cdot 3H_2O$/hr	7,200	19,000	48,000	96,000	144,000
Rotating drum-					
Ft^2 @ 12#/ft^2/hr.	600	1580	4,000	8,000	12,000
	\$265 M	\$490 M	\$880 M	\$1.36 MM	\$1.75 MM
Causticizer:					
gallons/hr.	22,000	58,000	145,000	290,000	440,000
includes agitator	\$37,000	\$50,000	\$110,000	\$215,000	\$325,000
Filter for Gypsum:					
# $CaSO_4 \cdot 2H_2O$/hr	8,000	20,000	50,000	100,000	150,000
Ft^2@ 13#/hr/ft^2	615	1540	3850	7,700	11,600
Rotary Drum	\$265 M	\$490 M	\$880 M	\$1.36 MM	\$1.75 MM

CAPITAL: (Continued)

For #/hour K_2SO_4 regeneration rates of:-

	7,500	20,000	50,000	100,000	150,000
Total bare module	\$4.61 MM	\$10. MM	\$19. MM	\$36. MM	\$50. MM
% of bare module = electrolytic	74%	72%	68.5%	69%	70%
Total Module cost =1.25xBare Mod.	\$5.7 MM	\$12.5 MM	\$24 MM	\$45 MM	\$63 MM

George T. Miller

FORMATE PROCESS

for seed regeneration from MHD

The process is to convert potassium sulfate to potassium formate as follows:-

$$K_2SO_4 + Ca(OH)_2 \rightleftharpoons 2KOH + CaSO_4$$
$$2KOH + 2CO \rightleftharpoons 2KCOOH$$

This process is discussed in the Un. of Tenn. Topical Report of Jan.4,1979 entitled "Evaluation of Alternative Seed Regeneration Processes Applicable to a Coal-Fired MHD Power Plant" by Matty et al . The appendix gives some detail and a flow-sheet. The process is based on a German patent (equivalent U.S. Pat. #2,030,082) and is reported to have been a commercial process. The following comments in the report are considered pertinent.

a) The CO utilization efficiency is reported as unknown.

b) The reaction kinetics are not well defined and thus the equipment size is not defined.

c) The economics are unknown.

d) Precipatation of potassium formate as discussed on Page 5 of the above report is questioned as is the filtration step before incineration. The formate is much too soluble (88.4%by wt.) to be in agreement with the statements and flow-sheet.

Based on personal experience with sodium formate from caustic and CO (a much more direct process), the following can be said:

a) Reaction rates are greatly dependant on partial pressure of the CO. Thus, high reaction rates and high utilization are favored by high gas pressure, high percentage CO in the supply gas and the number of countercurrent reaction stages used. Approximately 80% utilization is expected.

b) Experience with sodium formate would define minimum reactor size at 10# formate/ cubic foot/hour. This is based on a very effectively agitated reaction vessel at approximately 450psi and approximately 200°C. Use of a packed tower or bubble plate column is not recommended based on the experience with caustic of requiring over 100 times the reaction time as that required in a high shear agitated reactor for comparable results.

ASSUMPTIONS AND RECOMMENDATIONS:

In that three primary centers of economic concern are:-1) CO utilization, 2) reactor throughput and capital cost and 3) evaporation to recover the desired product, I recommend:

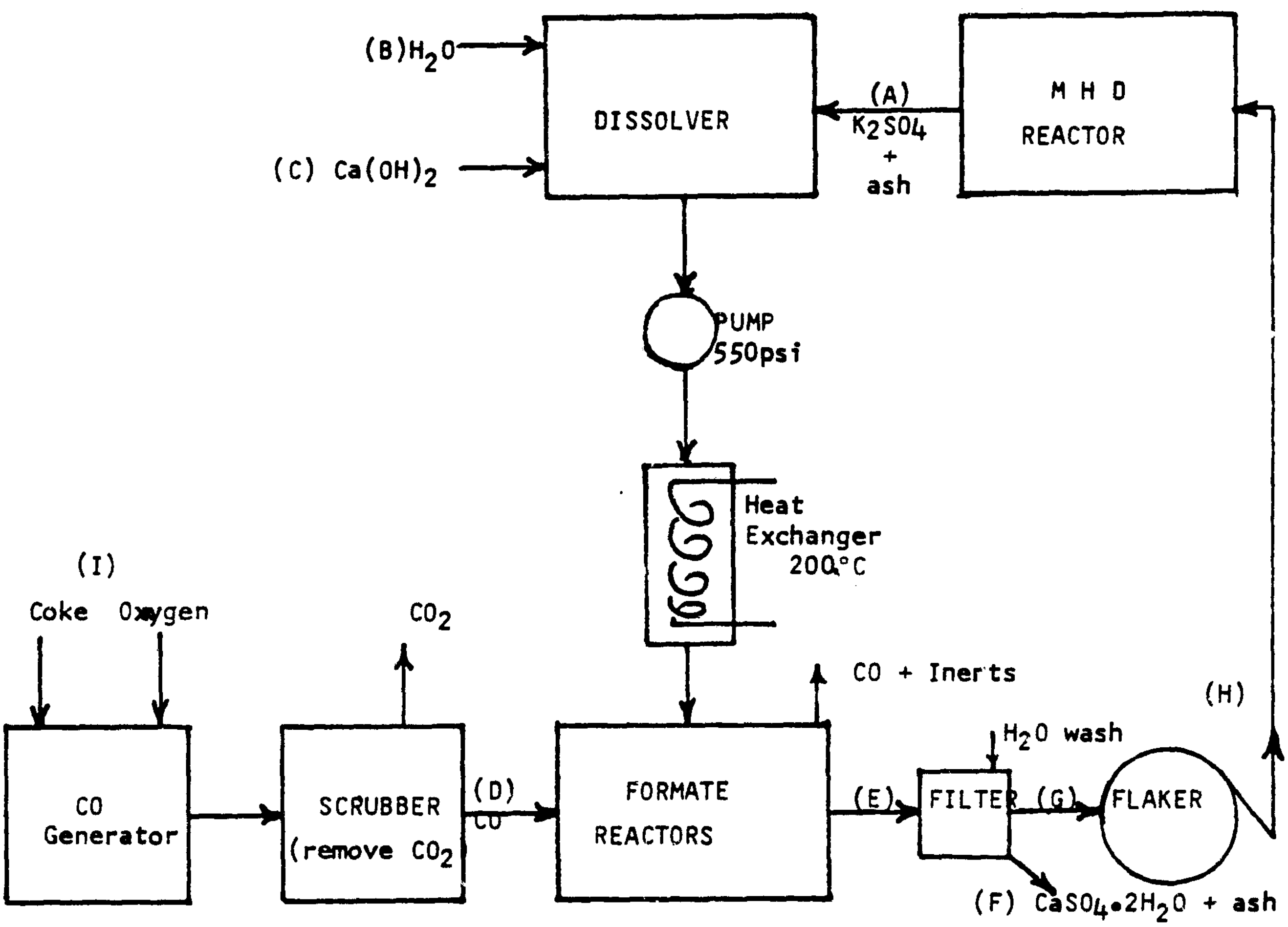
a) Use of a slurry of $Ca(OH)_2$ and K_2SO_4 (with attendant ash) in the Formate Reactors so that the product from the reactors, when filtered to remove $CaSO_4$ and ash, will yield an approximate 50-60% solution. This stream can then be evaporated on a 150°C drum flaker to produce a KCOOH + K_2SO_4 product for feed to the MHD plant. This eliminates the evaporators and decreases the size of the CO reactors.

b) Generate CO from the reaction between coke and pure oxygen to maintain a high CO concentration in the gas, thus greatly decreasing reactor investment,etc.

c) Use of high shear agitation in the CO reactors.

d) Reactors to be 'Monel' clad to minimize corrosion.

e) Minimum reactor size is based on 10# Formate/cu.ft./hr. The more realistic production capability is 3.3 # formate/cu.ft./hour and this will be used for investment computations.



Letter reference is to the flow-sheet on page-2. All weights are expressed in pounds per hour. Assume 20% excess K_2SO_4 is recycled through the formate process to drive the equilibrium in the formate reactors.

A_1	K_2SO_4 to be regenerated	7,500	20,000	50,000	100,000	150,000
A_2	Total K_2SO_4	9,000	24,000	60,000	120,000	180,000
B	H_2O	9,000	24,000	60,000	120,000	180,000
C	$Ca(OH)_2$	3,200	8,500	21,300	42,000	64,000
D	CO (80%utilized)	3,000	8,000	20,000	40,000	60,000
E	K_2SO_4	1,500	4,000	10,000	20,000	30,000
	KCOOH	7,240	19,300	48,300	96,500	144,000
	$CaSO_4$	5,900	15,600	39,000	78,000	117,000
F	$CaSO_4 \cdot 2H_2O$	7,500	20,000	50,000	100,000	150,000
	Ash (5%of K_2SO_4)	450	1,000	2,500	5,000	7,500
G	K_2SO_4	1,500	4,000	10,000	20,000	30,000
	KCOOH	7,240	19,300	48,300	96,500	144,000
	H_2O	9,000	24,000	60,000	120,000	180,000
H	K_2SO_4	1,500	4,000	10,000	20,000	30,000
	KCOOH	7,240	19,300	48,300	96,500	144,000
I	O_2(80%CO+20%CO_2)	2,600	6,900	17,000	35,000	52,000
	Coke @ 10%ash	1,900	4,800	12,000	24,000	36,000

DEBITS :- (per hour)

O_2 @ $20/ton	$26	$69	$170	$350	$520
$Ca(OH)_2$(@$35/ton CaO)	$43	$113	$282	$564	$848
MWe for agitation reactors,dissolvers	.37	.746	1.19	2.39	3.58
(@ 20 mil/KWH)	$7.40	$14.90	$23.80	$47.80	$71.60

No debit taken for preheat or water evaporation from the flaker in that there is more than enough heat from the CO generator for these applications.

CAPITAL :-

Capital for individual items expressed as 'bare module' which includes all costs of equipment, installation,piping,steel,electrical,concrete, labor, site and similar direct and indirect costs but does not include contingency and contractor fees. The contingency and contractor fees will be added to the sum of the individual components to get the total module cost, the module being the seed regeneration plant.
Note that 'M' = thoudsands and 'MM' = millions. Costs of major equipment were computed from 1968 cost data as recorded in Chemical Engineering, March 24,1969 page 114 and adjusted to 1979 costs using a factor of 2.08.

For #/hour K_2SO_4 regeneration rates of:	7,500	20,000	50,000	100,000	150,000
Dissolvers:					
3 hour retention					
10% leeboard-steel					
+agitation -gallons	7,000	18,000	44,000	87,000	130,000
	\$37,000	\$50,000	\$110,000	\$215,000	\$325,000
Pumps:					
550 psi for gal/hr	2,200	5,750	14,400	28,800	43,200
	\$20,000	\$33,500	\$80,000	\$160,000	\$240,000
Formate Reactors:					
for #/hr formate	7,240	19,300	48,300	96,500	144,800
Ft^3 @ 10#/hr/ft^3	724	1,930	4,830	9,650	14,480
Used a max. size					
of 8'D.x8'H	\$0.36MM	\$0.92MM	\$1.73MM	\$3.5MM	\$5.2MM
Filters:					
#$CaSO_4 \cdot 2H_2O$+ash/hr	8,000	21,000	53,000	105,000	160,000
Ft^2 @ 13#/hr/ft^2	615	1,615	4,100	8,200	12,300
	\$0.266MM	\$0.49MM	\$0.88MM	\$1.36MM	\$1.76MM
Flaker:(evaporator)					
# KCOOH+K_2SO_4/ hr.	9,000	24,000	60,000	120,000	180,000
Ft^2 @ 8 #/hr/ft^2	1,150	3,000	7,500	15,000	22,500
	\$0.391MM	\$0.723MM	\$1.30MM	\$2.03MM	\$2.63MM
CO Generator:					
#CO/hour	3,000	8,000	20,000	40,000	60,000
BTU of combustion	79 MM	210 MM	525 MM	1,050 MM	1,575 MM
	\$0.661MM	\$1.37MM	\$3.15MM	\$5.65MM	\$7.85MM
Scrubber(CO_2 removal)					
CO_2 to be removed					
#/hr. @80%CO,20%CO_2	750	2,000	5,000	10,000	15,000
Ft^3 @ 4 ft^3/#CO_2/hr.	3,000	8,000	20,000	40,000	60,000
Bubble plates @					
2'height between pl.	\$0.173MM	\$0.5 MM	\$1.5MM	\$2.5MM	\$4.0MM

CAPITAL: (continued)

For #/hour K_2SO_4 regeneration rates of:

	7,500	20,000	50,000	100,000	150,000
Total Bare Module	$1.9 MM	$4.1 MM	$8.8 MM	$15.4MM	$22.0MM
% of bare module = reactors	19%	22.5%	20%	23%	24%
Total Module cost =1.25xbare module	$2.4MM	$5.1MM	$11.MM	$20.MM	$27.5MM

George T. Miller

APPENDIX F

COST BASIS FOR COMBUSTION AND HEAT EXCHANGER SUBSYSTEMS

APPENDIX F

COST BASIS FOR COMBUSTOR AND HEAT EXCHANGER SUBSYSTEMS

Summaries of direct costs for the HTAH gasifiers/combustors, HTAH assembly and MHD combustors for Base Case 1, 2, and 3 are given in Tables F-1, F-2, and F-3, respectively. The figures quoted are direct cost F.O.B.; they include the costs of auxiliary components, instrumentation and control, but do not include the costs of vessel installation and erection except for the items listed below.

1. The HTAH costs include the direct costs and installation of: ducting, piping, insulation, valves, auxiliary components, flue gas recirculation fans, instrumentation and control, expansion joints and gas burners located within the battery limits of the HTAH.

2. The HTAH costs do not include the direct costs and installation of: main air compressors, expansion turbines, HTAH Gasifiers/Combustors, interface ducting between HTAH gasifiers and the HTAH assembly.

3. CAPFB costs include: 3 gasifier/sulfate generator vessels, 3 gasifier cyclones, 3 sulfate generator cyclones, and ducting between the gasifier and cyclones.

4. CAPFB costs do not include field erection of the 18.3 ft I.D. gasifier vessels and the following costs:

   Gasifier Field Erection
   (47 in. I.D.) Post Cyclone Product Gas Duct, $850/ft (3 each)
   (26 in. I.D.) Post Cyclone Off gas Duct, $650/ft (3 each)
   (105 in. I.D.) Main Manifold Duct, $2050/ft (1 each).

SOA Fixed Bed Gasifier

Cost data for the conventional fixed bed gasifier was obtained from McDowell Wellman, Cleveland, Ohio[1], a manufacturer of fixed bed gasifiers. In large production quantities, the F.O.B. cost per 10 ft I.D. gasifier, including clean-up cyclones, is $300,000. Additional operating and cost data of the SOA fixed bed gasifier are given in Table F-4.

Table F-1.

Cost Summary for HTAH Gasifier, HTAH Assembly and MHD Combustor

Base Case #1

Date 6 March 1979
Revision # 2

Case	Component	Component Description	Number of Vessels	Total Weight	Space Envelope L X W X H	Direct Cost
#				lbs $\times 10^6$	ftXftXft	$ X 10^6
1.0	HTAH Gasifier	Wellman Galusha	35	4.900	288x56x70	10.50
	HTAH	SOA HX	14	88.000	350x200x110	131.63
	MHD Combustor	2 Stage Cyclone	8 + 1	0.592	60x60x50	10.66
1.1	HTAH Gasifier	Wellman Galusha	30	4.200	240x56x70	9.00
	HTAH	SOA HX	14	88.000	350x200x110	131.63
	MHD Combustor	2 Stage Cyclone	8 + 1	0.408	50x50x50	9.80
1.2	HTAH Gasifier	Wellman Galusha	42	5.880	336x56x70	12.60
	HTAH	SOA HX	17	106.860	420x200x110	159.83
	MHD Combustor	2 Stage Cyclone	12 + 1	0.614	75x50x50	14.28
1.3	HTAH Gasifier	Wellman Galusha	35	4.900	288x56x70	10.50
	HTAH	SOA HX	14	88.000	350x200x110	131.63
	MHD Combustor	2 Stage Cyclone	8 + 1	0.407	50x50x50	9.90
1.4	HTAH Gasifier	Wellman Galusha	35	4.900	288x56x70	10.50
	HTAH	SOA HX	14	88.00	350x200x110	131.63
	MHD Combustor	1 Stage Vortex	6	0.684	54x25x30	6.84
1.5	HTAH Gasifier	Wellman Galusha	30	4.200	240x56x70	9.00
	HTAH	SOA HX	14	88.000	350x200x110	131.63
	MHD Combustor	1 Stage Vortex	6	0.548	50x25x30	5.75

Table F-2.

Cost Summary for HTAH Gasifier, HTAH Assembly and MHD Combustor

Base Case #2

Case #	Component	Component Description	Number of Vessels	Total Weight lbs x 10^6	Space Envelope L X W X H ftXftXftX	Direct Cost \$ X 10^6
2.0	HTAH Gasifier	2 Stage Cyclone Gasifier	5	0.273	35x35x40	5.90
	HTAH	Advanced, Cold Bottom HX	19	45.058	310x99x89	136.21
	MHD Combustor	1 Stage Vortex	6	0.862	40x34x30	6.69
2.0(a)	HTAH Gasifier	S^3PMB	15 + (2)	9.350	155x54x70	34.00
	HTAH	Advanced, Cold Bottom HX	18	47.214	311x100x93	142.73
	MHD Combustor	S^3PMB + 2nd Stage Coal Combustor	1	0.250	15x15x20	1.00
2.0(b)	HTAH Gasifier	S^3PMB	10 + (2)	6.600	110x54x70	24.00
	HTAH	Advanced, Cold Bottom HX	14	36.050	272x87x93	108.98
	MHD Combustor	S^3PMB + SPMB	14 + (2) + 1	8.800	144x54x70	33.00
2.1	HTAH Gasifier	2 Stage Cyclone Gasifier	4	0.218	35x35x40	4.72
	HTAH	Advanced, Cold Bottom HX	17	41.497	290x93x89	125.44
	MHD Combustor	1 Stage Vortex	6	0.862	40x34x30	6.69
2.2	HTAH Gasifier	2 Stage Cyclone Gasifier	4	0.312	30x30x40	5.14
	HTAH	Advanced, Hot Bottom NX	17	29.427	296x95x80	88.96
	MHD Combustor	2 Stage Cyclone Combustor	12 + 1	0.724	75x50x50	14.50
2.4	HTAH Gasifier	2 Stage Cyclone Gasifier	5	0.258	40x40x40	5.55
	HTAH	Advanced, Cold Bottom HX	18	41.743	302x97x87	126.19
	MHD Combustor	2 Stage Cyclone Combustor	9 + 1	0.573	60x50x50	11.24
2.5	Same as Case 2.0 Except for Pressure (Cs Seed)		4	0.227	30x30x40	4.92
			16	37.696	286x91x88	133.96
			6	0.568	39x34x30	5.80

Table F-2. (Continued)

Cost Summary for HTAH Gasifier, HTAH Assembly and MHD Combustor

Base Case #2

Case	Component	Component Description	Number of Vessels	Total Weight	Space Envelope L x W x H	Direct Cost
#				lbs x 10^6	ftXftXft	$ x 10^6
2.6	Same as Case 2.0 Except for Pressure (Supersonic Channel)		6	0.353	50x35x40	7.63
			25	63.025	360x115x93	190.525
			8	0.880	50x34x30	8.98
2.7	Same as Case 2.0 Except for Pressure (8T Magnet)		5	0.258	40x40x40	5.58
			17	41.452	298x95x89	125.308
			6	0.908	41x35x30	6.29
2.10	HTAH Gasifier	2 Stage Cyclone Gasifier	3	0.163	20x35x40	3.54
	HTAH	Advanced, Cold Bottom HX	10	25.138	228x73x92	75.99
	MHD Combustor	1 Stage Vortex	3	0.431	20x34x30	3.34
2.11	HTAH Gasifier	2 Stage Cyclone Gasifier	4	0.218	30x35x40	4.72
			14	33.105	263x84x91	100.08
	HTAH	Advanced, Cold Bottom HX				
	MHD Combustor	1 Stage Vortex	5	0.431	30x34x30	5.57
2.12	HTAH Gasifier	Air Cooled Vortex Burners	90	4.500	Included with HTAH	18.00
	HTAH	Advanced HX, atm Reheat	18	121.295	386x220x120	181.46
	MHD Combustor	1 Stage Vortex	6	0.600	40x30x30	6.12
2.15	HTAH Gasifier	2 Stage Cyclone Gasifier	5	0.273	35x35x40	5.90
	HTAH	Advanced, Cold Bottom HX	17	41.452	298x95x89	125.31
	MHD Combustor	1 Stage Vortex	12	0.431	80x34x30	13.38
2.16	HTAH Gasifier	2 Stage Cyclone Gasifier	4	0.170	25x25x40	3.67
	HTAH	Advanced, Hot Bottom HX	17	22.100	257x82x80	66.81
	MHD Combustor	1 Stage Vortex	6	0.542	38x34x30	5.53
2.17	HTAH Gasifier	CAPFB	3	1.410	50x50x70	2.98
	HTAH	Advanced, Cold Bottom HX		41.452	298x95x89	125.31
	MHD Combustor	1 Stage Vortex	6	0.862	40x34x30	6.69

Table F-3.

Cost Summary for MHD Combustor

Base Case #3

Case	Component	Component Description	Number of Vessels	Total Weight	Space Envelope L X W X H	Direct Cost
#				lbs x 10^6	ftXftXftX	\$ x 10^6
3.0	MHD Combustor	1 Stage Vortex	8	0.864	72x25x30	8.880
3.1	MHD Combustor	1 Stage Vortex	8	0.727	66x25x30	7.440
3.2	MHD Combustor	2 Stage Cyclone Gasifier	10 + 1	0.647	60x60x50	12.83
3.4	MHD Combustor	1 Stage Vortex	8	0.880	75x25x30	8.97
3.5	MHD Combustor	1 Stage Vortex	6	0.622	54x25x30	6.35

Table F-4. Reference Data for Cost Scaling

COMPONENT	UNITS	FIXED BED GASIFIER	SOA HTAH	TWO STAGE CYCLONE GASIFIER/COMBUSTOR	SINGLE STAGE VORTEX COMBUSTOR	ADVANCED HTAH PRESSURIZED REHEAT	ADVANCED HTAH atm PREHEAT	CAPFB	SPMB/S^{3}PMB
Pressure	atm	1.25	1.15/8.0	7.4	6.0	7.0	1.15/9.	8.0	6.15
$\dot{m}_{coal}$(total)	Kg/s	37.01/28.53	-	24.8	7.32	-	-	48.11	4.71
$\dot{m}_{air}$(total)	Kg/s	-	458.0	-	-	268.1	529.4	-	-
ΔTair	°F	-	2100	-	-	2400	2384	-	-
ΔHair	MW/Kg/s	-	625.2	-	-	422.5	826.9	-	-
Capacity per vessel	Kg/s	1.07/0.945	32.714	6.2	7.32	22.34	[illegible].603	16.036	4.71
Characteristic Size D X L	ft x ft	10 x 12	34 x 70	5.7 x 14	4.0 x 13.6	14 x 37	34 x 8[illegible]	20.3 x 52	13.1 x 15
Cost per Unit	$ X 10^6	0.300	9.402	1.125 (1.000)	1.020	7.451	10.674	1.063	2.000
Weight per Unit (with Fuel)	16 x 10^6	0.140	6.286	0.052	0.100	2.464	7.135	0.503	0.550
Space Envelope L X W X H	ft x ft x ft	16 x 28 x 70 (each gasifier)	350 x 200 x 110 (14 vessels)	25 x 50 x 40 (4 primary Gasifiers)	18 x 10 x 30 (each Combustor)	250 x 80 x 90 (Bank of 12)	386 x 220 x 120 (Bank of 17)	54 x 54 x 70 (Bank of 3)	20 x 28 x 70 (each Gasifier)
Vessels Required	No.	35/30	14	4	1	12	17	3	1
Case #	No.	1.0	1.0	-	AVCO ETF	-	2.12	-	-
Remarks		MR/I6 Data from McDowell-Wellman	atm Reheat/ Pressurized Blowdown, Scaled McKee Data	Estimated Cost for 2nd Stage Combustor. Scaled from GE ETF	Data from AVCO ETF Report	Scaled from GE ETF Report	Scaled from Case 1.0	Data from FWDC	Data extrapolated from several sources

SOA HTAH

Direct costs for the SOA HTAH were obtained via extrapolation of cost data provided in the GE ETF Study[2]. In scaling to the commercial size plant, it was assumed that 60% of the direct cost scaled linearily with preheat air delivery requirements while the remaining 40% scaled by the square root of the delivered mass flow rate. This scaling was adapted on the assumptions that refractory costs, which constitute 60% of the HTAH costs, will scale linearily with plant size while the remaining 40% will accrue benefits from economies of scale and only scale as the square root of air mass flow rate. Changes in operating pressure during blowdown in Base Case 1 were not reflected in vessel size changes since it is assumed the vessel designs are pressure drop limited during the reheat cycle.

Two Stage High Slag Rejection Gasifier/Combustor

Data for costing this component was obtained from extrapolation of data developed during the GE ETF Study[2]. The direct cost for a clustered gasifier (nominally 2 each 5.7 ft I.D. primary gasifiers per cluster) is $2,225,000. This translates into a cost of 1,125,000 per gasifier with an accounting for component manifolding. At a design pressure of 7.4 atm each gasifier has a rated capacity of 6.2 Kg/s. Reference cost data for this component are given in Table F-4. The capacity of the gasifier was scaled linearly with pressure for operating pressures differing from the design condition. The diameter of the gasifier was then adjusted to accommodate a suitable number of gasifiers for acceptable clustering arrangements. For these cases requiring a second stage gas phase combustor, $1,000,000 was allocated to account for the cost of the second stage gas phase combustor and required

hot manifold ducting.

Single Stage Vortex Combustor

Cost data for the single stage vortex combustor was obtained from the AVCO ETF report[3]. Scale-up of this component was assumed to be via the use of ETF scale combustor modules. The cost scaling rationale for this combustor is the same as for the two-stage cyclone combustor concept except that an accounting for a second stage gas phase combustor was not required.

Advanced HTAH (Pressurized Reheat/Pressurized Blowdown)

Costs for the advanced HTAH were obtained by scaling cost and performance data from the ETF[2] design (7 ft O.D. vessels) to a commercial size reference design (14 ft O.D. vessels). The costs for the reference design were obtained from the ETF size HTAH by assuming 60% of the material cost scaled linearly with air capacity while the remaining 40% scaled with the square root of air capacity. Table F-4 provides the cost data for this reference design. Since the capital investment for the HTAH is significant, it was deemed desirable to scale the vessel size (and cost) as a function of delivered air flow, air temperature rise, .d operating pressure. The air mass flow capacity per vessel, $\dot{m}$, through a given HTAH vessel can be scaled by fixing the percent pressure drop across the HTAH and is scaled as follows:

$$\frac{\dot{m}(2)}{\dot{m}(1)} = \left|\frac{D(2)}{D(1)}\right|^{5/2} \left|\frac{P(2)}{P(1)}\right| \left|\frac{\Delta T(1)}{\Delta T(2)}\right|^{1/2}, \qquad (1)$$

where the (2) refers to perturbed conditions, and the (1) refers to the reference operating conditions; the D, P and ΔT refer to the matrix diameter, operating pressure and air temperature increase through the heat exchanger,

respectively. Operating conditions for perturbation cases were determined after channel/generator calculations optimized net electrical output from the MHD flow train. Having determined the air mass flow capacity per vessel, the approximate number of vessels needed was calculated. The number of vessels required was then rounded off and adjusted mass flow capacities and vessel sizes calculated. The specific weight adjustments from the reference case were then proportioned to D^2L, with D being the HTAH diameter and L being the matrix bed height which was assumed to be proportional to ΔT. The total cost for the system was then calculated from the total weight by ascribing a fixed cost per pound. The space envelope for the HTAH assembly was determined from the number and size of the vessels relative to the reference design case.

Advanced HTAH (Atmospheric Reheat/Pressurized Blowdown)

Costing for the HTAH in Case 2.12 (3000 F delivered air) developed as a perturbation to the concepts developed for Base Case 1 (2700 F delivered air). The air capacity per vessel was scaled inversely proportional to the square root of the air temperature increase while the weight of each vessel was scaled in proportion; i.e.,

$$\frac{\dot{m}(2)}{\dot{m}(1)} = \left|\frac{\Delta T(1)}{\Delta T(2)}\right|^{1/2} \text{, and} \tag{10}$$

$$\frac{Wt(2)}{Wt(1)} = \frac{\Delta T(2)}{\Delta T(1)} \tag{11}$$

A summary of the reference cost data for this HTAH assembly is given in Table F-4.

CAPFB

Cost data for the CAPFB in Case 2.17 were obtained by scaling information provided by the Foster Wheeler Development Corporation for 20.25 ft. I.D. gasifiers designed for operation at 8 atm pressure. The reference cost data for this component are given in Table F-4.

The diameter of the vessel was scaled according to the following relationship

$$D(2) = D(1) \left[\frac{\dot{m}(2)}{\dot{m}(1)}\right]^{1/2} \left[\frac{P(1)}{P(2)}\right]^{1/2} , \qquad (12)$$

which assumes the superficial gas velocity is held constant, while the weight and cost were assumed to scale as the surface area, which in this case implies a linear scaling with diameter since the vessel height does not change appreciably.

The cost for field construction and erection of the gasifier is not included as part of the cost quoted in Table F-2. Field erection costs for this type component could vary between 100% and 400% of the F.O.B. material cost.

Slagging Pressurized Moving Bed Gasifier (SPMB/S^3PMB)

Cost estimates for the SPMB were obtained from cost data published in a recent EPRI Report[4] and via personnal communications with EPRI[5] and Lurgi of America[6]. The estimates direct cost for a 4 m SPMB is $2,000,000 each. Erection and installation costs are estimated at 200% of the F.O.B. material cost. Data from the above referenced EPRI report suggested the use of spare gasifier

Spare gasifiers are included in the costing for Case 2.0(a) and 2.0(b); the spare units are identified by parentheses in Table F-2. Costs for the S^3PMB and SPMB were assumed to be the same since the construction of the gasifiers would likely be similar in nature. Reference performance and cost data for the SPMB are given in Table F-4. Adjustments to cost were not adjusted for operating pressures differing from the pressure because of uncertainties associated with the base cost as well as vessel capacity.

References-Appendix F

1. Personal Communication (1978, 1979) Wallace Hamilton, Senior Consultant, McDowell-Wellman Company, Cleveland, Ohio.

2. Section 1.1, Reference 6.

3. Section 1.1, Reference 7.

4. Chandra, R., McElmurry, B., and Smelser, S., (1978) "Economics of Fuel Gas From Coal-An Up Date Including the British Gas Corporation's Slagging Gasifier," Prepared by Flow Engineering for Electric Power Research Institute, AF-782, May 1978.

5. Personal Communications (1979) Nevil Halt, Clean Fossil Fuels Dept., Electric Power Research Institute, Palo Alto, California.

6. Personal Communications (1979) Ted Pollaert, Lurgi of America; New York City, New York.

APPENDIX G

SAMPLE COSTING PROCEDURE, CASE 3.5

APPENDIX G

COSTING PROCEDURE, CASE 3.5

G.1.0 DETERMINATION OF CAPITAL COSTS (Refer to Table G-1)

1. Determine Major Equipment Cost, BOP Material Costs and Installation Costs for each account using the input parameters for Case 3.5, and the scaling factors derived from the appropriate Bechtel cost estimate. (In this case, the scaling factors derived from Bechtel's estimate of Case 3.0 were used).

2. Estimate Indirect Field Costs for each account by taking 75% of the installation cost determined for that acocunt.

3. Field Construction Cost is the sum of the total cost for each account or the sume of the totals for each cost category:

   Field Construction Cost = 894.8

4. Engineering Services is 15% of the sum of BOP Materials, Installation and Indirect Costs:

   Engineering Services = 0.15 (141.5 + 134.9 + 101.2)
   = 0.15 (377.6)
   = 56.6

5. Contingency is 10% of the sum of the conventional plant costs and the applicable portion of their engineering services cost plus 20% of the MHD Topping Cycle costs (Account No. 317) and the applicable portion of the engineering services costs.

   Contingency (MHD Components) = 0.20 [541.2 + 0.15 (32.5 + 50.0 + 37.5)]
   = 111.84

   Contingency (Conventional Plant) = 0.10 [(894.8 - 541.2) + 0.15 (377.6 - 120.0)]
   = 39.224

   Total Contingency = 151.1

6. Oxygen Plant Cost, if applicable, is determined using the estimate of the Lotepro Oxygen Plant cost, scaled up to the appropriate capacity, (See Table G-2)

7. Escalation and Interest During Construction (E & IDC) was done assuming a 6-1/2 year construction period, applying the E&I factor of 1.679 (see Table G-4) to the sum of the Field Construction, Engineering Services, Contingency and Oxygen Plant Costs, and de-escalating to current dollars (6-1/2 years at 6-1/2 %/year). This figure is the Total Estimated Construction Cost (TECC) expressed in current dollars. Subtracting the initial sum from the TECC gives the E&IDC cost.

Field Construction	894.8
Engineering Services	56.6
Contingency	151.1
Oxygen Plant	85.6
Sum	1188.1

Escalation and Interest Multiplier

$$\frac{1.679}{(1.065)^{6.5}} = 1.11501$$

Total Estimated Construction Cost

1188.1 x 1.11501 = 1324.7

Escalation and Interest During Construction

1324.7 - 1188.1 = 136.1

G.2 DETERMINATION OF COST OF ELECTRICITY (Refer to Table G-3)

1. Capital Cost:

   a. Fixed Charge Rate = 18%/yr
   b. Plant Capacity Factor = 0.65
   c. Total Estimated Construction Cost = $\$1324.7 \times 10^6$
   d. Plant Capacity = 1118.7 MWe

$$\text{Capital COE} = \frac{1324.7 \times 10^6\ (0.18)}{(1118.7)\ (0.65)\ (8760\ \text{hrs/yr})}$$

2. Fuel Cost:

   a. Cost Levelization Factor = 1.882
      (Based on 6-1/2% Inflation Rate for 30 Year Plant Life)
   b. Current Fuel Cost = $\$1.05/10^6$ BTU
   c. Plant Efficiency = 44.07%

$$\text{Fuel COE} = \frac{1.05\ (1.882)\ (3.413\ \text{BTU/watt-hr}}{0.4407}$$

$$= 15.30\ \text{mills/KWhr}$$

3. Operating and Maintenance Cost

Data/Assumptions:

Plant Capacity = 1118.7 MWe
Seed Reprocessing O&M Cost = 0.76 Mills/KWhr
Cost Levelization Factor = 1.882
MHD Generator Cost = $\$17.5 \times 10^6$
MHD Generator Service Life = 10,000 hr.
Plant Life = 30 years
Hours Per Year = 8760
Capacity Factor = 0.65

a. Maintenance Cost:

Fixed Maintenance Cost - ECAS II data (except MHD Generator) escalated from mid 1975 to mid 1978 at 8% per year:

1.46 mills/KWhr

MHD Generator, 10,000 hours between overhaul, 1/2 cost of new generator at each overhaul:

$$\frac{\$17.5 \times 10^6 \times 0.5}{1118.7 \times 10{,}000} = 0.09\ \text{mills/KWhr}$$

Spare MHD Generator amortized over 30 year plant life

$$\frac{17.5 \times 10^6}{1118.7 \times 8760 \times 30 \times 0.65} = 0.78\ \text{mills/KWhr}$$

Total Maintenance Cost = 2.33 mills/KWhr

b. <u>Operating Cost</u>:

Crew of 130 (based on ECAS II) at $25,000/year

$$\frac{130 \times \$25{,}000}{1118.7 \times 0.65 \times 8760} \; 0.51 \text{ mills/KWhr}$$

c. <u>Operating Consumables</u>

Conventional Operating Costs (Non-MHD) based on ECAS II data escalated to mid 1978

$$\frac{\$794{,}000}{1118.7 \times 0.65 \times 8760} \; 0.12 \text{ mills/KWhr}$$

Seed Reprocessing (Table 3.9-3)

0.76 mills/KWHR

d. <u>Levelized Total Operating and Maintenance Cost</u>:

3.72 x 1.882 = 7.00 mills/KWhr

Table G-1. Plant Capital Cost Estimate Summary
Open Cycle MHD

Case 3.5

		Million $				
Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	35.8	31.1	23.3	90.2
312	Boiler Plant Equipment	70.2	16.1	21.3	16.0	123.6
314	Turbogenerator Units	25.8	27.4	13.7	10.3	77.2
315	Accessory Electrical Equipment	-	19.9	18.0	13.5	51.4
316	Miscellaneous Power Plant Equipment	-	3.2	0.7	0.5	4.4
317	MHD Topping Cycle	421.2	32.5	50.0	37.5	541.
350	Transmission Plant	-	6.6	0.1	0.1	6.8
	FIELD CONSTRUCTION COST	517.2	141.5	134.9	101.2	894.8
	Engineering Services					56.6
	Contingency					151.1
	Oxygen Plant					85.6
	Other Costs - Escalation and Interest During Construction					136.6
	TOTAL ESTIMATED CONSTRUCTED COST					1324.7

Table G-2. O_2 Plant Costs

Cost basis: \$50.58 x 10^6 per 6087 tons/day (197.8 KWhr/ton equivalent pure O_2)

CASE NO.	P_{O_2} (MW)	TONS/DAY	COST (\$ x 10^6)
1.0	32.7	3968	33.0
1.1	32.7	3968	33.0
1.2	0	-	-
1.3	32.7	3968	33.0
1.4	32.7	3968	33.0
1.4a	32.7	3968	33.0
3.0	84.9	10,301	85.6
3.1	84.7	10,277	85.4
3.2	84.9	10,301	85.6
3.4	86.3	10,471	87.0
3.5	84.9	10,301	85.6

Note: O_2 plant is assumed to be delivered as a Turnkey installation hence no installation, indirect, engineering or contingency costs are included.

Table G-3. Cost Summary for Case 3.5

CAPITAL COST: ($ x 10^6)	
MAJOR EQUIPMENT COST	[illegible]94.2
BOP MATERIAL COST	141.5
INSTALLATION COST	134.9
INDIRECT FIELD COST	101.2
FIELD CONSTRUCTION COST	871.8
ENGINEERING SERVICES	56.6
CONTINGENCY	146.4
OXYGEN PLANT	85.6
ESCALATION AND INTEREST DURING CONSTRUCTION	133.5
TOTAL CAPITAL COST	1293.9
PLANT OUTPUT (MWe)	1118.7
CAPITAL COST ($/kWe)	1156.6
COST OF ELECTRICITY: (Mills/KW/hr)	
CAPITAL COST	36.56
FUEL COST	15.30
OPERATION AND MAINTENANCE	7.00
TOTAL COE	58.86
OVERALL PLANT EFFICIENCY (%)	44.07

Table G-4. Escalation and Interest Cost Factors

[Escalation + Interest = Total. Annual rates: escalation, 6.5 percent; interest, 10 percent.]

Time from start of design to powerplant completion, yr	Escalation	Interest on obligated funds	Total
	Cost factor		
0	1.000	1.000	1.000
.5	1.018	1.022	1.040
1.0	1.037	1.044	1.081
1.5	1.056	1.069	1.125
2.0	1.076	1.094	1.170
2.5	1.096	1.122	1.218
3.0	1.116	1.151	1.267
3.5	1.137	1.182	1.319
4.0	1.158	1.214	1.372
4.5	1.179	1.249	1.428
5.0	1.202	1.285	1.487
5.5	1.224	1.324	1.548
6.0	1.247	1.365	1.612
6.5	1.270	1.409	1.679
7.0	1.294	1.454	1.748
7.5	1.319	1.503	1.822
8.0	1.344	1.554	1.898
8.5	1.369	1.609	1.978
9.0	1.395	1.666	2.061
9.5	1.422	1.726	2.148
10.0	1.449	1.790	2.239

APPENDIX H

PLANT CAPITAL COST ESTIMATE SUMMARIES

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 1.0

(Primary Change From Reference Case: Reference)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	--	35.3	31.2	23.4	89.9
312	Boiler Plant Equipment	85.8	18.1	21.5	16.2	141.6
314	Turbogenerator Units	27.9	31.5	15.6	11.7	86.7
315	Accessory Electrical Equipment	--	14.4	16.7	12.5	43.6
316	Miscellaneous Power Plant Equipment	--	3.4	0.7	0.5	4.6
317	MHD Topping Cycle	377.5	55.3	51.8	38.9	523.5
350	Transmission Plant	--	6.4	0.1	0.1	6.6
	FIELD CONSTRUCTION COST	491.2	164.4	137.6	103.3	896.5
	Engineering Services					60.8
	Contingency					150.3
	Oxygen Plant					33.0
	Other Costs - Escalation and Interest during Construction					131.1
	TOTAL ESTIMATED CONSTRUCTED COST					1271.7

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 1.1

(Primary Change From Reference Case: I6)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	35.5	31.1	23.3	89.9
312	Boiler Plant Equipment	85.4	17.9	21.3	16.0	140.6
314	Turbogenerator Units	27.3	31.0	15.4	11.5	85.2
315	Accessory Electrical Equipment	-	14.2	16.6	12.5	43.3
316	Miscellaneous Power Plant Equipment	-	3.4	0.7	0.5	4.6
317	MHD Topping Cycle	375.5	51.9	49.5	37.1	514.0
350	Transmission Plant	-	6.5	0.1	0.1	6.7
	FIELD CONSTRUCTION COST	488.2	160.4	134.7	101.1	884.3
	Engineering Services					59.4
	Contingency					147.9
	Oxygen Plant					33.0
	Other Costs - Escalation and Interest during Construction					129.3
	TOTAL ESTIMATED CONSTRUCTED COST					1253.9

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 1.2

(Primary Change From Reference Case: Air)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	34.9	31.3	23.4	89.6
312	Boiler Plant Equipment	88.7	19.4	22.8	17.1	148.0
314	Turbogenerator Units	29.2	32.8	16.2	12.2	90.4
315	Accessory Electrical Equipment	-	12.1	16.2	12.2	40.5
316	Miscellaneous Power Plant Equipment	-	3.5	0.7	0.5	4.7
317	MHD Topping Cycle	410.0	61.6	57.9	43.4	572.9
350	Transmission Plant	-	6.5	0.1	0.1	6.7
	FIELD CONSTRUCTION COST	527.9	170.8	145.2	108.9	952.8
	Engineering Services					63.7
	Contingency					161.4
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					135.4
	TOTAL ESTIMATED CONSTRUCTED COST					1313.3

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 1.3

(Primary Change From Reference Case: 7T)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	35.8	31.4	23.6	90.8
312	Boiler Plant Equipment	83.9	17.8	21.1	15.8	138.6
314	Turbogenerator Units	27.0	30.7	15.2	11.4	84.3
315	Accessory Electrical Equipment	-	14.5	16.7	12.5	43.7
316	Miscellaneous Power Plant Equipment	-	3.3	0.7	0.5	4.5
317	MHD Topping Cycle	465.6	56.6	53.1	39.8	615.1
350	Transmission Plant	-	6.5	0.1	0.1	6.7
	FIELD CONSTRUCTION COST	576.5	165.2	138.3	103.7	983.7
	Engineering Services					61.1
	Contingency					168.2
	Oxygen Plant					33.0
	Other Costs - Escalation and Interest during Construction					143.3
	TOTAL ESTIMATED CONSTRUCTED COST					1389.3

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 1.4

(Primary Change From Reference Case: Single Stage)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	35.4	31.3	23.5	90.2
312	Boiler Plant Equipment	86.9	18.1	21.7	16.2	142.9
314	Turbogenerator Units	27.9	31.3	15.5	11.6	86.3
315	Accessory Electrical Equipment	-	14.4	16.7	12.5	43.6
316	Miscellaneous Power Plant Equipment	-	3.4	0.7	0.5	4.6
317	MHD Topping Cycle	374.9	55.3	51.8	38.8	520.8
350	Transmission Plant	-	6.5	0.1	0.1	6.7
	FIELD CONSTRUCTION COST	489.7	164.4	137.8	103.2	895.1
	Engineering Services					60.8
	Contingency					149.9
	Oxygen Plant					33.0
	Other Costs - Escalation and Interest during Construction					130.9
	TOTAL ESTIMATED CONSTRUCTED COST					1269.7

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 1.4a

(Primary Change From Reference Case: Slag Rejection)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	35.1	31.2	23.4	89.7
312	Boiler Plant Equipment	85.6	18.1	21.5	16.1	141.3
314	Turbogenerator Units	28.6	31.8	15.8	11.8	88.0
315	Accessory Electrical Equipment	-	14.3	16.7	12.5	43.5
316	Miscellaneous Power Plant Equipment	-	3.4	0.7	0.5	4.6
317	MHD Topping Cycle	370.2	55.1	51.6	38.7	515.6
350	Transmission Plant	-	6.4	0.1	0.1	6.6
	FIELD CONSTRUCTION COST	484.4	164.2	137.6	103.1	889.3
	Engineering Services					60.7
	Contingency					148.8
	Oxygen Plant					33.0
	Other Costs - Escalation and Interest during Construction					130.1
	TOTAL ESTIMATED CONSTRUCTED COST					1261.9

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.0

(Primary Change From Reference Case: Reference)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	--	35.3	31.6	23.7	90.6
312	Boiler Plant Equipment	83.3	14.3	13.4	10.1	121.1
314	Turbogenerator Units	28.5	29.6	14.8	11.1	84.0
315	Accessory Electrical Equipment	--	13.2	16.5	12.4	42.1
316	Miscellaneous Power Plant Equipment	--	2.9	0.6	0.4	3.9
317	MHD Topping Cycle	395.3	50.3	44.8	33.6	524.0
350	Transmission Plant	--	7.1	0.1	0.1	7.3
	FIELD CONSTRUCTION COST	507.1	152.7	121.8	91.4	873.0
	Engineering Services					54.9
	Contingency					147.1
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					123.6
	TOTAL ESTIMATED CONSTRUCTED COST					1198.6

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.0s

(Primary Change From Reference Case: Slag)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	--	34.9	31.4	23.6	89.9
312	Boiler Plant Equipment	84.7	14.5	13.6	10.2	123.0
314	Turbogenerator Units	29.3	30.1	15.0	11.3	85.7
315	Accessory Electrical Equipment	--	13.2	16.5	12.4	42.1
316	Miscellaneous Power Plant Equipment	--	2.9	0.6	0.5	4.0
317	MHD Topping Cycle	392.8	50.4	44.9	33.7	521.8
350	Transmission Plant	--	7.1	0.1	0.1	7.3
	FIELD CONSTRUCTION COST	506.8	153.1	122.1	91.8	873.8
	Engineering Services					55.1
	Contingency					147.0
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					123.7
	TOTAL ESTIMATED CONSTRUCTED COST					1199.6

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.0a

(Primary Change From Reference Case: S^3PMB + Coal)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	--	32.7	29.5	22.1	84.3
312	Boiler Plant Equipment	75.6	13.8	12.6	9.4	111.4
314	Turbogenerator Units	26.2	27.2	13.6	10.2	77.2
315	Accessory Electrical Equipment	--	13.2	16.5	12.4	42.1
316	Miscellaneous Power Plant Equipment	--	2.6	0.5	0.4	3.5
317	MHD Topping Cycle	418.2	51.8	47.5	35.6	553.1
350	Transmission Plant	--	6.6	0.1	0.1	6.8
	FIELD CONSTRUCTION COST	520.0	147.9	120.3	90.2	878.4
	Engineering Services					53.8
	Contingency					150.5
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					124.5
	TOTAL ESTIMATED CONSTRUCTED COST					1207.2

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.0b

(Primary Change From Reference Case: S^3PMB + SPMB)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	--	32.5	29.3	22.0	83.8
312	Boiler Plant Equipment	68.7	13.8	12.5	9.4	104.4
314	Turbogenerator Units	26.1	27.0	13.5	10.1	76.7
315	Accessory Electrical Equipment	--	13.1	16.5	12.3	41.9
316	Miscellaneous Power Plant Equipment	--	2.6	0.5	0.4	3.5
317	MHD Topping Cycle	407.8	57.3	56.5	42.3	563.9
350	Transmission Plant	--	6.5	0.1	0.1	6.7
	FIELD CONSTRUCTION COST	502.6	152.8	128.9	96.6	880.9
	Engineering Services					56.7
	Contingency					152.5
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					125.4
	TOTAL ESTIMATED CONSTRUCTED COST					1215.5

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.1

(Primary Change From Reference Case: I6)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	--	35.6	31.6	23.7	90.9
312	Boiler Plant Equipment	84.1	14.3	13.4	10.1	121.9
314	Turbogenerator Units	28.7	30.1	15.0	11.3	85.1
315	Accessory Electrical Equipment	--	13.1	16.5	12.3	41.9
316	Miscellaneous Power Plant Equipment	--	2.9	0.6	0.5	4.0
317	MHD Topping Cycle	398.8	48.2	43.1	32.3	522.4
350	Transmission Plant	--	7.2	0.1	0.1	7.4
	FIELD CONSTRUCTION COST	511.6	151.4	120.3	90.3	873.6
	Engineering Services					54.3
	Contingency					146.9
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					123.6
	TOTAL ESTIMATED CONSTRUCTED COST					1198.4

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.2

(Primary Change From Reference Case: Two-Stage Cyclone, Hot Bottom HTAH)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	--	33.3	29.7	22.2	85.2
312	Boiler Plant Equipment	76.2	13.6	12.5	9.4	111.7
314	Turbogenerator Units	26.0	27.3	13.6	10.2	77.1
315	Accessory Electrical Equipment	--	13.1	16.5	12.3	41.9
316	Miscellaneous Power Plant Equipment	--	2.7	0.5	0.4	3.6
317	MHD Topping Cycle	360.4	46.8	41.7	31.3	480.2
350	Transmission Plant	--	6.6	0.1	0.1	6.8
	FIELD CONSTRUCTION COST	462.6	143.4	114.6	85.9	806.5
	Engineering Services					51.6
	Contingency					135.6
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					114.3
	TOTAL ESTIMATED CONSTRUCTED COST					1108.0

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.2a

(Primary Change From Reference Case: Two-Stage Cyclone)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	--	35.1	31.4	23.6	90.1
312	Boiler Plant Equipment	84.1	14.4	13.5	10.1	122.1
314	Turbogenerator Units	28.8	29.9	14.9	11.2	84.8
315	Accessory Electrical Equipment	--	13.2	16.5	12.4	42.1
316	Miscellaneous Power Plant Equipment	--	2.9	0.6	0.5	4.0
317	MHD Topping Cycle	400.8	50.3	44.8	33.6	529.5
350	Transmission Plant	--	7.0	0.1	0.1	7.2
	FIELD CONSTRUCTION COST	513.7	152.8	121.8	91.5	879.8
	Engineering Services					54.9
	Contingency					148.4
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					124.6
	TOTAL ESTIMATED CONSTRUCTED COST					1207.7

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.4

(Primary Change From Reference Case: NASA Specification)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	--	35.1	31.4	23.6	90.1
312	Boiler Plant Equipment	83.8	14.4	13.5	10.1	121.8
314	Turbogenerator Units	28.0	29.8	14.9	11.2	83.9
315	Accessory Electrical Equipment	--	12.0	16.2	12.2	40.4
316	Miscellaneous Power Plant Equipment	--	2.9	0.6	0.4	3.9
317	MHD Topping Cycle	384.6	49.7	44.1	33.1	511.5
350	Transmission Plant	--	7.0	0.1	0.1	7.2
	FIELD CONSTRUCTION COST	496.4	150.9	120.8	90.7	858.8
	Engineering Services					54.4
	Contingency					144.4
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					121.6
	TOTAL ESTIMATED CONSTRUCTED COST					1179.2

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.4a

(Primary Change From Reference Case: GE Recalculation)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	35.1	31.5	23.6	90.2
312	Boiler Plant Equipment	83.9	14.4	13.5	10.1	121.9
314	Turbogenerator Units	29.0	29.8	14.9	11.2	84.9
315	Accessory Electrical Equipment	-	13.3	16.5	12.4	42.2
316	Miscellaneous Power Plant Equipment	-	2.9	0.6	0.5	4.0
317	MHD Topping Cycle	383.9	50.0	44.3	33.2	511.4
350	Transmission Plant	-	7.1	0.1	0.1	7.3
	FIELD CONSTRUCTION COST	496.8	152.6	121.4	91.1	861.9
	Engineering Services					54.8
	Contingency					144.7
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					122.1
	TOTAL ESTIMATED CONSTRUCTED COST					1183.5

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.5

(Primary Change From Reference Case: Cs Seed)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	36.1	31.9	23.9	91.9
312	Boiler Plant Equipment	80.7	14.0	13.1	9.8	117.6
314	Turbogenerator Units	27.1	28.6	14.3	10.7	80.7
315	Accessory Electrical Equipment	-	13.1	16.5	12.3	41.9
316	Miscellaneous Power Plant Equipment	-	2.8	0.6	0.4	3.8
317	MHD Topping Cycle	375.5	49.4	43.5	32.7	501.1
350	Transmission Plant	-	7.2	0.1	0.1	7.4
	FIELD CONSTRUCTION COST	483.3	151.2	120.0	89.9	844.4
	Engineering Services					54.2
	Contingency					141:9
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					119.7
	TOTAL ESTIMATED CONSTRUCTED COST					1160.2

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.6

(Primary Change From Reference Case: Super Sonic)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	34.7	31.4	23.6	89.7
312	Boiler Plant Equipment	85.4	14.5	13.7	10.3	123.9
314	Turbogenerator Units	30.2	30.4	15.2	11.4	87.2
315	Accessory Electrical Equipment	-	13.4	16.5	12.4	42.3
316	Miscellaneous Power Plant Equipment	-	3.0	0.6	0.5	4.1
317	MHD Topping Cycle	485.9	52.3	47.7	35.8	621.7
350	Transmission Plant	-	7.1	0.1	0.1	7.3
	FIELD CONSTRUCTION COST	601.5	155.4	125.2	94.1	976.2
	Engineering Services					56.2
	Contingency					167.4
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					138.0
	TOTAL ESTIMATED CONSTRUCTED COST					1337.8

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.7

(Primary Change From Reference Case: 8 T)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	36.2	32.0	24.0	92.2
312	Boiler Plant Equipment	80.7	14.0	13.1	9.8	117.6
314	Turbogenerator Units	27.5	28.6	14.3	10.7	81.1
315	Accessory Electrical Equipment	-	13.2	16.5	12.4	42.1
316	Miscellaneous Power Plant Equipment	-	2.8	0.6	0.4	3.8
317	MHD Topping Cycle	568.4	51.0	45.9	34.5	699.8
350	Transmission Plant	-	7.3	0.1	0.1	7.5
	FIELD CONSTRUCTION COST	676.6	153.1	122.5	91.9	1044.1
	Engineering Services					55.1
	Contingency					181.9
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					147.3
	TOTAL ESTIMATED CONSTRUCTED COST					1428.4

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.10

(Primary Change From Reference Case: 1500 MWt)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	18.9	17.9	13.4	50.2
312	Boiler Plant Equipment	47.0	8.6	8.0	6.0	69.6
314	Turbogenerator Units	18.6	16.7	8.3	6.3	49.9
315	Accessory Electrical Equipment	-	10.7	15.9	11.9	38.5
316	Miscellaneous Power Plant Equipment	-	1.6	0.3	0.3	2.2
317	MHD Topping Cycle	225.0	28.5	28.4	21.3	303.2
350	Transmission Plant	-	3.7	0.1	0.1	3.9
	FIELD CONSTRUCTION COST	290.6	88.7	78.9	59.3	517.5
	Engineering Services					34.0
	Contingency					86.6
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					73.4
	TOTAL ESTIMATED CONSTRUCTED COST					711.5

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.11

(Primary Change From Reference Case: 2000 MWt)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	25.2	23.2	17.4	65.8
312	Boiler Plant Equipment	61.1	10.8	10.1	7.6	89.6
314	Turbogenerator Units	22.6	21.6	10.8	8.1	63.1
315	Accessory Electrical Equipment	-	11.7	16.1	12.1	39.9
316	Miscellaneous Power Plant Equipment	-	2.1	0.4	0.3	2.8
317	MHD Topping Cycle	299.2	36.9	34.8	26.1	397.0
350	Transmission Plant	-	5.0	0.1	0.1	5.2
	FIELD CONSTRUCTION COST	382.9	113.3	95.5	71.7	663.4
	Engineering Services					42.1
	Contingency					111.7
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					94.0
	TOTAL ESTIMATED CONSTRUCTED COST					911.2

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.11a

(Primary Change From Reference Case: Slag Rejection)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	24.9	23.0	17.3	65.2
312	Boiler Plant Equipment	62.0	10.9	10.2	7.7	90.8
314	Turbogenerator Units	23.3	22.0	11.0	8.3	64.6
315	Accessory Electrical Equipment	-	11.7	16.1	12.1	39.9
316	Miscellaneous Power Plant Equipment	-	2.1	0.4	0.3	2.8
317	MHD Topping Cycle	297.1	37.1	34.9	26.2	395.3
350	Transmission Plant	-	5.0	0.1	0.1	5.2
	FIELD CONSTRUCTION COST	382.4	113.7	95.7	72.0	663.8
	Engineering Services					42.2
	Contingency					111.6
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					94.0
	TOTAL ESTIMATED CONSTRUCTED COST					911.6

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.12

(Primary Change From Reference Case: Atm Air Heater Combustor)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	34.9	31.1	23.3	89.3
312	Boiler Plant Equipment	82.3	14.1	13.2	9.9	119.5
314	Turbogenerator Units	27.6	29.1	14.5	10.9	82.1
315	Accessory Electrical Equipment	-	14.7	16.8	12.6	44.1
316	Miscellaneous Power Plant Equipment	-	2.8	0.6	0.4	3.8
317	MHD Topping Cycle	442.1	57.5	57.1	42.8	599.5
350	Transmission Plant	-	6.9	0.1	0.1	7.1
	FIELD CONSTRUCTION COST	552.0	160.0	133.4	100.0	945.4
	Engineering Services					59.0
	Contingency					162.8
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					134.2
	TOTAL ESTIMATED CONSTRUCTED COST					1301.4

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.15

(Primary Change From Reference Case: Dual Flow Train)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	50.1	42.5	31.9	124.5
312	Boiler Plant Equipment	83.3	14.3	13.4	10.1	121.1
314	Turbogenerator Units	28.5	29.6	14.8	11.1	84.0
315	Accessory Electrical Equipment	-	13.2	16.5	12.4	42.1
316	Miscellaneous Power Plant Equipment	-	2.9	0.6	0.4	3.9
317	MHD Topping Cycle	559.9	78.1	65.4	49.1	752.5
350	Transmission Plant	-	7.1	0.1	0.1	7.3
	FIELD CONSTRUCTION COST	671.7	195.3	153.3	115.1	1135.4
	Engineering Services					69.6
	Contingency					198.6
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					161.4
	TOTAL ESTIMATED CONSTRUCTED COST					1565.0

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.16

(Primary Change From Reference Case: Hot Bottom HTAH, 4 kV/m)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	34.4	30.2	22.7	87.3
312	Boiler Plant Equipment	73.7	13.3	12.2	9.1	108.3
314	Turbogenerator Units	24.5	26.3	13.2	9.9	73.9
315	Accessory Electrical Equipment	-	13.0	16.4	12.3	41.7
316	Miscellaneous Power Plant Equipment	-	2.6	0.5	0.4	3.5
317	MHD Topping Cycle	333.0	46.0	40.5	30.4	449.9
350	Transmission Plant	-	6.8	0.1	0.1	7.0
	FIELD CONSTRUCTION COST	431.2	142.4	113.1	84.9	771.6
	Engineering Services					51.1
	Contingency					129.0
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					109.5
	TOTAL ESTIMATED CONSTRUCTED COST					1061.2

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.16a

(Primary Change From Reference Case: Hot Bottom HTAH, 6 - 5 T)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	33.6	29.8	22.3	85.7
312	Boiler Plant Equipment	75.3	13.5	12.4	9.3	110.5
314	Turbogenerator Units	25.7	27.0	13.5	10.1	76.3
315	Accessory Electrical Equipment	-	13.1	16.5	12.3	41.9
316	Miscellaneous Power Plant Equipment	-	2.6	0.5	0.4	3.5
317	MHD Topping Cycle	329.6	46.1	40.6	30.4	446.7
350	Transmission Plant	-	6.7	0.1	0.1	6.9
	FIELD CONSTRUCTION COST	430.6	142.6	113.4	84.9	771.5
	Engineering Services					51.1
	Contingency					128.7
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					109.4
	TOTAL ESTIMATED CONSTRUCTED COST					1060.7

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.16b

(Primary Change From Reference Case: Cold Bottom HTAH, 4 kV/m)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	36.0	31.8	23.8	91.6
312	Boiler Plant Equipment	81.0	14.1	13.1	9.9	118.1
314	Turbogenerator Units	27.0	28.7	14.4	10.8	80.9
315	Accessory Electrical Equipment	-	13.1	16.5	12.3	41.9
316	Miscellaneous Power Plant Equipment	-	2.8	0.6	0.4	3.8
317	MHD Topping Cycle	398.1	50.0	44.7	33.5	526.3
350	Transmission Plant	-	7.2	0.1	0.1	7.4
	FIELD CONSTRUCTION COST	506.1	151.9	121.2	90.8	870.0
	Engineering Services					54.6
	Contingency					147.0
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					123.2
	TOTAL ESTIMATED CONSTRUCTED COST					1194.8

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.17

(Primary Change From Reference Case: CAPFB Air Heater Combustor)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	36.1	32.3	24.2	92.6
312	Boiler Plant Equipment	80.9	14.5	13.6	10.2	119.2
314	Turbogenerator Units	29.3	30.7	15.4	11.5	86.9
315	Accessory Electrical Equipment	-	13.3	16.5	12.4	42.2
316	Miscellaneous Power Plant Equipment	-	3.0	0.6	0.5	4.1
317	MHD Topping Cycle	381.9	51.0	44.9	33.7	511.5
350	Transmission Plant	-	7.2	0.1	0.1	7.4
	FIELD CONSTRUCTION COST	492.1	155.8	123.4	92.6	863.9
	Engineering Services					55.8
	Contingency					145.1
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					122.5
	TOTAL ESTIMATED CONSTRUCTED COST					1187.3

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 2.18

(Primary Change From Reference Case: 1300 F Air and Flue Gas to Preheat Combustor)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	33.5	29.8	22.3	85.6
312	Boiler Plant Equipment	76.2	13.5	12.5	9.4	111.6
314	Turbogenerator Units	25.8	26.8	13.4	10.1	76.1
315	Accessory Electrical Equipment	-	13.2	16.5	12.4	42.1
316	Miscellaneous Power Plant Equipment	-	2.6	0.5	0.4	3.5
317	MHD Topping Cycle	389.4	48.3	44.0	33.0	514.7
350	Transmission Plant	-	6.7	0.1	0.1	6.9
	FIELD CONSTRUCTION COST	491.4	144.6	116.8	87.7	840.5
	Engineering Services					52.4
	Contingency					142.6
	Oxygen Plant					NA
	Other Costs - Escalation and Interest during Construction					119.1
	TOTAL ESTIMATED CONSTRUCTED COST					1154.6

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 3.0

(Primary Change From Reference Case: Ref.)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	34.5	30.6	22.9	88.0
312	Boiler Plant Equipment	73.9	16.6	22.1	16.6	129.2
314	Turbogenerator Units	28.2	28.9	14.5	10.9	82.5
315	Accessory Electrical Equipment	-	19.9	18.0	13.5	51.4
316	Miscellaneous Power Plant Equipment	-	3.4	0.7	0.5	4.6
317	MHD Topping Cycle	224.2	30.6	37.1	27.8	319.7
350	Transmission Plant	-	6.4	0.1	0.1	6.6
	FIELD CONSTRUCTION COST	326.3	140.3	123.1	92.3	682.0
	Engineering Services					53.4
	Contingency					106.9
	Oxygen Plant					85.6
	Other Costs - Escalation and Interest during Construction					106.7
	TOTAL ESTIMATED CONSTRUCTED COST					1034.6

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 3.1

(Primary Change From Reference Case: I6)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	35.0	30.6	23.0	88.6
312	Boiler Plant Equipment	71.8	16.2	21.5	16.2	125.7
314	Turbogenerator Units	26.8	28.1	14.1	10.6	79.6
315	Accessory Electrical Equipment	-	19.9	18.0	13.5	51.4
316	Miscellaneous Power Plant Equipment	-	3.3	0.7	0.5	4.5
317	MHD Topping Cycle	252.0	29.2	35.3	26.5	343.0
350	Transmission Plant	-	6.5	0.1	0.1	6.7
	FIELD CONSTRUCTION COST	350.6	138.2	120.3	90.4	699.5
	Engineering Services					52.4
	Contingency					110.8
	Oxygen Plant					85.4
	Other Costs - Escalation and Interest during Construction					109.0
	TOTAL ESTIMATED CONSTRUCTED COST					1057.1

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 3.2

(Primary Change From Reference Case: Two-Stage Cyclone)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	34.8	30.7	23.0	88.5
312	Boiler Plant Equipment	72.9	16.4	21.9	16.4	127.6
314	Turbogenerator Units	27.4	28.5	14.3	10.7	80.9
315	Accessory Electrical Equipment	-	19.9	18.0	13.5	51.4
316	Miscellaneous Power Plant Equipment	-	3.3	0.7	0.5	4.5
317	MHD Topping Cycle	233.3	30.5	36.4	27.3	327.5
350	Transmission Plant	-	6.4	0.1	0.1	6.6
	FIELD CONSTRUCTION COST	333.6	139.8	122.1	91.5	687.0
	Engineering Services					53.0
	Contingency					108.1
	Oxygen Plant					85.6
	Other Costs - Escalation and Interest during Construction					107.4
	TOTAL ESTIMATED CONSTRUCTED COST					1041.1

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 3.4

(Primary Change From Reference Case: 1100 F Air)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	34.8	31.0	23.2	89.0
312	Boiler Plant Equipment	76.0	17.0	22.6	16.9	1 2.5
314	Turbogenerator Units	28.9	29.7	14.9	11.2	84.7
315	Accessory Electrical Equipment	-	20.0	18.0	13.5	51.5
316	Miscellaneous Power Plant Equipment	-	3.5	0.7	0.5	4.7
317	MHD Topping Cycle	226.3	31.0	36.9	27.7	321.9
350	Transmission Plant	-	6.5	0.1	0.1	6.7
	FIELD CONSTRUCTION COST	331.2	142.5	124.2	93.1	691.0
	Engineering Services					53.9
	Contingency					108.1
	Oxygen Plant					87.0
	Other Costs - Escalation and Interest during Construction					108.1
	TOTAL ESTIMATED CONSTRUCTED COST					1048.1

PLANT CAPITAL COST ESTIMATE SUMMARY

(Mid 1978 Dollars x 10^6)

Case 3.5

(Primary Change From Reference Case: 8T)

Account No.	Account Description	Major Equipment	BOP	Inst. Cost	Indir. Cost	Total Cost
311	Structures and Improvement	-	35.8	31.1	23.3	90.2
312	Boiler Plant Equipment	70.2	16.1	21.3	16.0	123.6
314	Turbogenerator Units	25.8	27.4	13.7	10.3	77.2
315	Accessory Electrical Equipment	-	19.9	18.0	13.5	51.4
316	Miscellaneous Power Plant Equipment	-	3.2	0.7	0.5	4.4
317	MHD Topping Cycle	398.2	32.5	50.0	37.5	518.2
350	Transmission Plant	-	6.6	0.1	0.1	6.8
	FIELD CONSTRUCTION COST	494.2	141.5	134.9	101.2	871.8
	Engineering Services					56.6
	Contingency					146.4
	Oxygen Plant					85.6
	Other Costs - Escalation and Interest during Construction					133.5
	TOTAL ESTIMATED CONSTRUCTED COST					1293.9

APPENDIX I

SCALING OF MAGNET SIZE AND COST: DETAILS AND EXAMPLE

APPENDIX I

SCALING OF MAGNET SIZE AND COST: DETAILS AND EXAMPLE

This brief effort was based on the scale-up of the AVCO BL6-P1 design, Figure I-1, since that was one of the few design concepts scaled for base load size at the time of the study. The BL6 configuration was assumed for all cases and materials and current density were unchanged except for the superconductor for the 8 Tesla case. Summary tables and the general procedure for estimating magnet size and cost were given in Section 3.4. Additional details and an example are presented here. Table I-1 lists basic size data for the reference magnet and the four cases considered.

For magnets of the size of interest for PSPEC, the cost of superstructure can be expected to be one of the greatest single cost items. Superstructure weight and cost are influenced by many design parameters, as shown in Figure I-2. For the main section of the magnet and for the same material, the ring girder weight can be expected to scale as the product of magnet length, girder cross-section area and mean radius from warm bore axis to the ring girder. The selected girder can be expected to have a cross-section area which is scaled by some power less than unity (typically about 0.7) of the required section modulus. The section modulus of the ring girder, in turn, can be expected to scale approximately as the square of the product of magnetic field and magnet radius. When scaling from a given design, however, it must be recognized that that design may have a superstructure which is not optimally designed over the entire length of the magnet.

The mass of the conductor and the substructure for the main section of the magnet are expected to scale as the product of the magnet length, the build of the winding, and the RMS radius from the magnet center axis to the winding. The build ratio is determined by the ampere turns required to produce the required field for a warm bore of a given diameter.

The outer vacuum jacket cylinder walls are expected to scale as the product of the magnet length and the outside diameter of the magnet structure. The end sections of the vacuum jacket can be expected to scale as the square of that outer diameter.

Figure I-3 shows a numerical example, for Case 2.2, of the scaling procedure used. Note that in one of the other cases (Case 2.10), the principle items of mass were unchanged from the BL6-P1 reference even though the magnet dimensions were significantly different. In this particular case the product of certain scaling factors was found to be unity, even though some of these individual factors were greater than or smaller than unity.

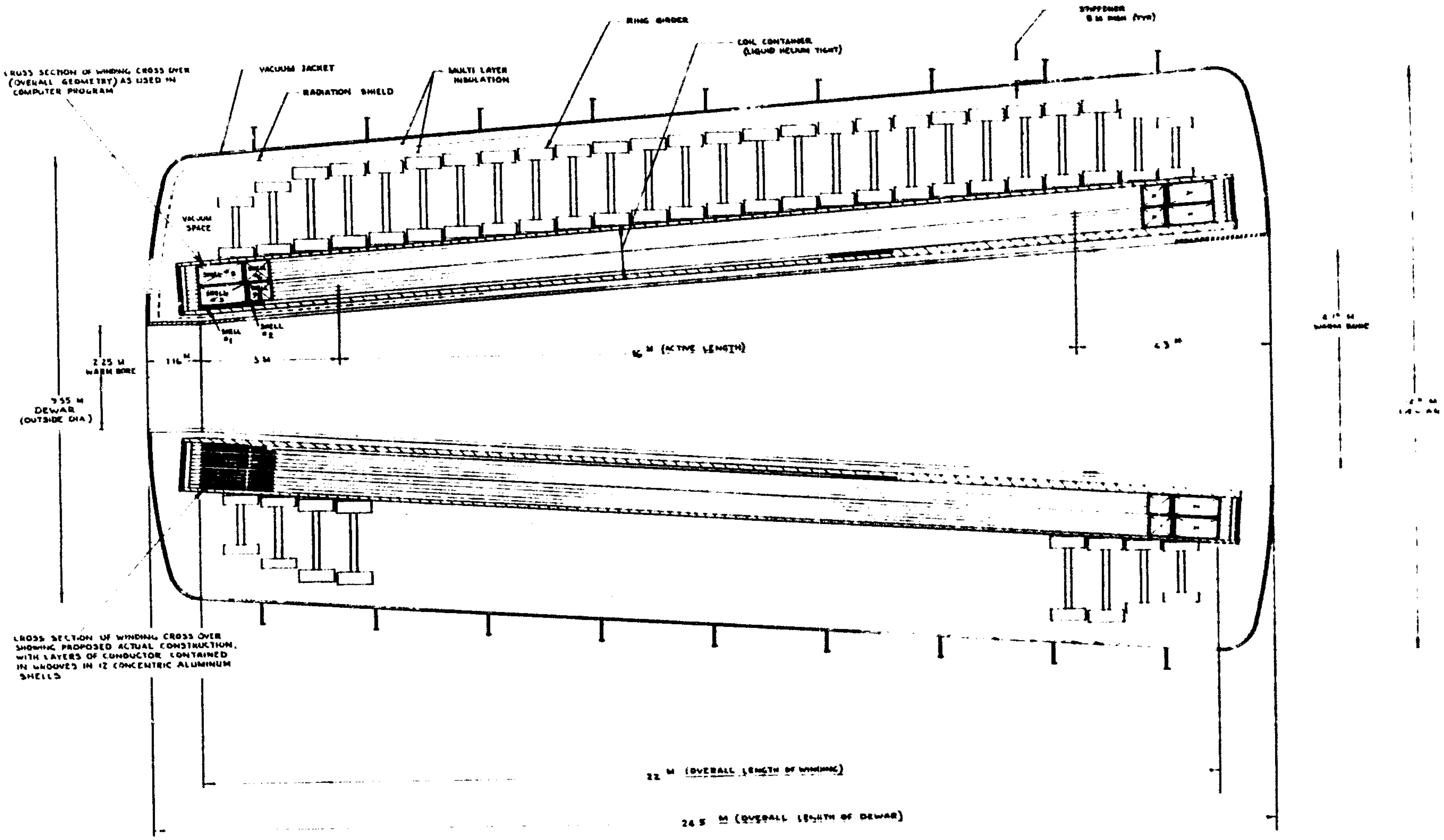


Figure I-1. Conceptual Design, 6 T Superconducting Circular Saddle Coil Magnet, Base-Load Size, Design BL6-1

Table I-1. Basic Size Data for PSPEC Magnets Compared to BL6-P1 Reference

Case	Output Power MWe	Active Length M	B_1	B_{L-1}	Mass Flow Rate (kg/sec)	Warm Bore Diameter		B x Diam	
						$\ell = 1$	$\ell = L-1$	$\ell = 1$	ℓ L-1
BL6-P1 (Reference)	-	16	5	3.5	-	2.69	4.84	16.1	16.9
2.2	1257	25	5	5	606	3.8	5.4	22.8	27
2.11	887	20	6	5	454	2.2	4.8	13.2	24
2.10	665	15	6	5	325	2.3	4.1	13.8	20.5
2.7	1293	25	3	7	606	3.8	5.4	30.4	37.8

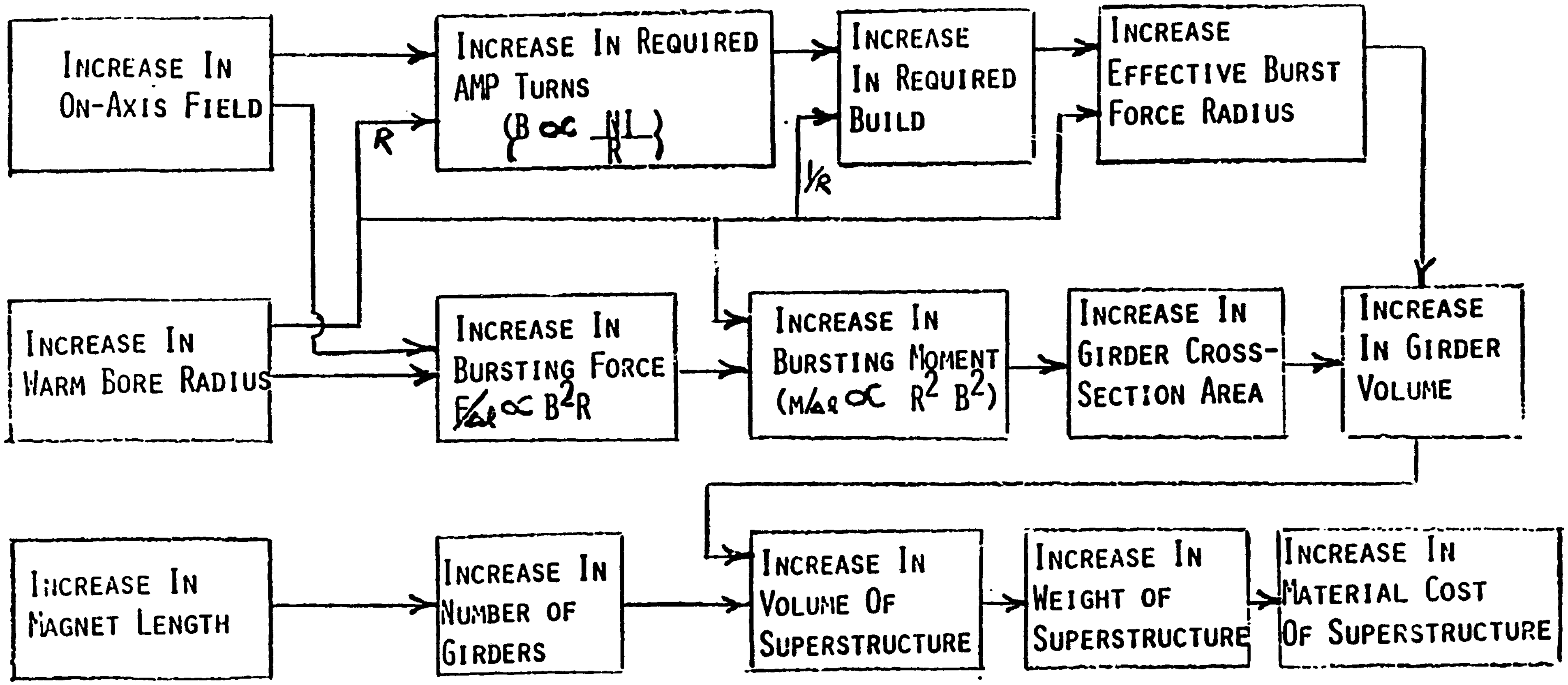


Figure I-2. A Rationale for Sup-rstructure Scaling

A SCALING EXAMPLE

CASE 2.2

BUILD:

INCREASE BUILD BY FIELD RATIO:

AT ENTRANCE: $\frac{6}{6} = 1$

AT EXIT: $\frac{5}{3.4} = 1.5$

DECIDE TO INCREASE BUILD BY 1.33 OR

0.9M x 1.33 = 1.2M

Figure I-3a.

SUPERSTRUCTURE

INCREASE GIRDER SECTION MODULUS BY B^2 R^2 RATIO

(R MEASURED TO CENTER OF WINDINGS):

B^2 R^2 RATIO

AT ENTRANCE	AT EXIT
$\frac{c^2}{6^2} \times \frac{3^2}{2.2^2} = 1.86$	$\frac{5^2}{3.4^2} \times \frac{3.9^2}{3.4^2} = 2.8$

BUT FOR BL6 - P1:

$$\frac{B^2\ R^2\ \text{(AT ENTRANCE)}}{B^2\ R^2\ \text{(AT EXIT)}} = \frac{6^2 \times 2.2^2}{3.4^2 \times 3.4^2} = 1.3$$

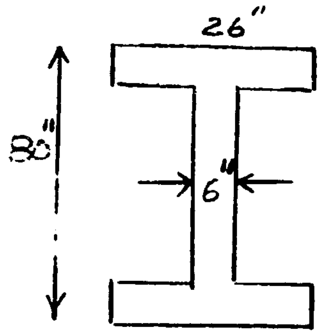
WHICH STATES FOR BL6 - P1: RING GIRDERS OVERDESIGNED AT EXIT BY FACTOR OF 1.3.

THEREFORE AT EXIT, SECTION MODULUS NEED BE INCREASED BY:

$$\frac{2.8}{1.3} = 2.15$$

Figure I-3b.

SECTION MODULUS (I/C) OF BL6 - P1



PER GIRDER: $I/C = 20{,}416\ \text{in}^3$ $\qquad A = 960\ \text{in}^2$

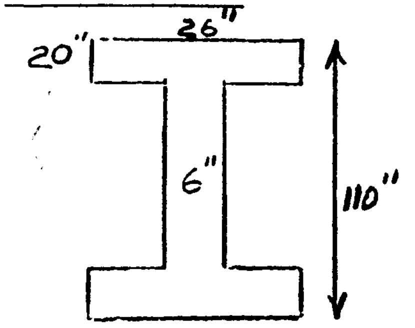
PER FOOT: $\dfrac{I/C}{\text{FT.}} = 9{,}422\ \text{in}^3/\text{FT}$ $\qquad A/\Delta\ell = 443\ \text{in}^2/\text{FT}$

THEREFORE CASE 2.2 NEEDS SECTION MODULUS OF:

$$9{,}422 \times 2.15 \approx 20{,}000\ \text{in}^3/\text{FT}$$

Figure I-3c.

SELECT GIRDER:



$I/C = 42{,}036 \text{ IN}^3$ A = 1460

$\frac{I/C}{\text{FT}} = 19{,}401 \text{ IN}^3/\text{FT}$ A/FT = 673

INCREASE IN SECTION MODULUS $= \frac{42{,}036}{20{,}416} = 2.1$

INCREASE IN AREA $= \frac{1460}{960} = 1.5$

Figure I-3d.

GIRDER WEIGHT

$$\left(\frac{10.5}{8.6}\right) \times (1.5) \times \left(\frac{25}{21}\right) = 2.3$$

RATIO* OF DIAMETER TO CENTER OF GIRDER	CROSS-SECTION AREA RATIO	LENGTH RATIO

CONDUCTOR AND SUBSTRUCTURE WEIGHT

$$\left(\frac{7}{5.9}\right) \times \left(\frac{1.2}{0.9}\right) \times \left(\frac{23}{19}\right) = 1.9$$

WINDING DIAMETER RATIO*	BUILD RATIO	COIL LENGTH RATIO

*RMS OF ENTRANCE AND EXIT VALUES BASED ON WARM BORE DIMENSIONS, PLUS DATA SCALED FROM FIGURE I-1.

Figure I-3e.

VACUUM JACKET

CYLINDER WALL VOLUME RATIO:

LENGTH RATIO x OUTSIDE DIAM. RATIO

$$\left(\frac{29.5}{24.5}\right) \times \left(\frac{15.2}{12.5}\right) = 1.5$$

LENGTH RATIO   OUTSIDE DIAM. RATIO

END WALLS VOLUME RATIO:

(OUTSIDE DIAMETER RATIO)^2

$$\left(\frac{(15.2)}{(12.5)}\right)^2 = 1.5$$

THEREFORE ASSUME TOTAL VACUUM JACKET VOLUME INCREASES BY 1.5

Figure I-3f.

APPENDIX J

NASA SPECIFIED CODE OF ACCOUNTS

APPENDIX J

NASA SPECIFIED CODE OF ACCOUNTS

Estimate Format, Guidelines and Modified FPC Code of Accounts

ETF ESTIMATE FORMAT

ACCT NO.	ACCOUNT DESCRIPTION	UNIT	QUANTITY	MATERIAL MJR COMP	COST BOP	INST COST	INDIR COST	CONTIN	TOTAL COST
	Direct Accounts (see attachment B)	xx	xxxx	xxxx	xxxx	xxxx	xxxx	xxxx	xxxx
	SUBTOTALS			xxxx	xxxx	xxxx	xxxx	xxxx	xxxx
	ENGINEERING SERVICES			-	-	xxxx	-	xxxx	xxxx
	OTHER COSTS			xxxx	xxxx	xxxx	xxxx	xxxx	xxxx
	TOTAL ESTIMATED COSTS			xxxx	xxxx	xxxx	xxxx	xxxx	xxxx

NOTES

1. Cost basis 1978 1/2 dollars.
2. Costs should be shown in thousands of dollars.
3. UNIT - Unit of measure for each account.
4. QUANTITY - Quantity for each account based on the unit of measure.
5. MATERIAL COST - Total material cost for each account.
   MJR COMP - Total major component material cost delivered to the site for each account. Major components are those major purchased items which are engineered, designed, fabricated, shipped, and in some cases erected by one supplier or manufacturer.
   BOP - Total balance of plant materials cost delivered to the site for each account. BOP materials are those items normally designed and purchased by an Architect - Engineer.
6. INST COST - Total direct installation cost for each account. Provide as part of the estimate backup a breakdown of those items included in the installation cost.
7. INDIR COST - Total indirect construction cost for each account. Provide as part of the estimate backup a breakdown of those items included in indirect construction cost and the procedure used in applying that cost to the estimate.
8. CONTIN - Total contingency cost for each account. Provide as part of the estimate backup the basis for the contingency costs and the procedure used in applying those costs to the estimate.
9. TOTAL COST - Total of all material, installation, indirect and contingency costs for each account.

10. ENGINEERING SERVICES - Total cost of all professional services. Provide as part of the estimate backup the basis for the engineering services cost and procedure used in applying that cost to the estimate.

11. OTHER COSTS - Total of other costs. Provide as part of the estimate backup a breakdown of those items included in other costs and the procedure in applying that cost to the estimate.

ETF CODE OF ACCOUNTS

ACCT NO.		DESCRIPTION
310		Land and Land Rights
311	(1)	Structures and Improvements
311.1		Improvements to Site
311.2		Main Building
311.3		Steam Turbine Building
311.4		Coal Bunker/Processing Area
311.5	(2)	Service Buildings
311.6	(3)	Other Buildings and Structures
312		Boiler Plant Equipment
312.1	(4)	Coal Handling and Processing
312.2	(5)	Slag and Ash Handling
312.3		Radiant Section
312.4	(6)	Steam Generator Sections
312.5	(7)	Effluent Control
312.6		Auxiliary Boiler Systems
312.7	(8)	Other Boiler Plant Systems
313	(9)	Engines and Engine Driven Generators
314		Turbogenerator Units
314.1		Steam Turbine Generator and Auxiliaries
314.2		Condenser and Auxiliaries
314.3	(10)	Circulating Water System
314.4	(11)	Steam Piping Systems
314.5		Other Turbine Plant Equipment
315	(12)	Accessory Electric Equipment
316	(13)	Miscellaneous Power Plant Equipment

317		MHD Topping Cycle Equipment
317.1	(14)	Combustion Equipment
317.2	(15)	MHD Generator Subsystem
317.3	(16)	Magnet Subsystem
317.4	(17)	Inverters and Electrode Control
317.5	(18)	Oxidizer Preheater Subsystem
317.6	(19)	Seed Subsystem
317.7	(20)	Other Major MHD Topping Cycle Support Subsystems
317.8		Miscellaneous MHD Topping Cycle Support Equipment
318	(21)	Research Equipment
319	(22)	Simulation Equipment
350	(23)	Transmission Plant

ETF CODE OF ACCOUNTS

GENERAL NOTES

A. The modified FPC "ETF Code of Accounts" and these notes should be used as a guide in developing and utilizing the final code of accounts for the ETF cost estimate.

B. These notes should be used as a guide in subdividing accounts and additional subaccounts included in order to more accurately define the cost of the plant. Individual subaccounts should be included for each major component or subsystem.

C. If the total cost of an account or subaccount is greater than 5% of the total estimated plant cost the account or subaccount should be subdivided into its next lower level of detail.

D. Individual component or subsystem foundations, structural steel supports, access platforms, etc. should not be included in account 311, but are to be included as part of the BOP material cost and installation cost for the component or subsystem.

E. All piping; ducting; and electrical, mechanical and instrumentation and control equipment within a subsystem should be included in the cost of the subsystem.

SPECIFIC NOTES

1. Subdivide the building accounts into subaccounts for each individual building or major building area. (Example: If subaccount 311.2 Main Building includes the heater, MHD, cryogenic system, invertor and control buildings provide separate subaccounts under 311.2 for each of these buildings.) All building services should be included in the cost of the building.

2. Includes all office, shop, warehouse and maintenance buildings. Include a subaccount for each building provided.

3. Includes all miscellaneous buildings such as the water treatment or seed system buildings. Include a subaccount for each building provided.

4. Includes all equipment from the initial coal unloading point up to and including coal storage prior to final preparation (final preparation includes drying and pulverizing).

5. Includes all equipment from the initial collection equipment up to the storage area.

6. Include subaccount for each major steam generator section (examples: superheater, reheater, economizer, etc.).

7. Includes all equipment from the steam generator outlet up to and including the chimney, with subaccounts for each seed and/or cleanup subsystem, and the chimney.

8. Include subaccounts for the condensate system, boiler feedwater system, condensate pumps, boiler feed pumps, boiler plant related water treatment equipment and the secondary air system.

9. Do not use account 313.

10. Include a subaccount for the cooling towers.

11. Include a subaccount for each of the major steam systems (examples: main steam, hot and cold reheat steam, extraction steam and bypass steam systems).

12. Includes all accessory electric equipment such as the equipment from the MHD power conditioning equipment and steam turbine generator up to the main transformers, emergency or standby equipment, and wire and cable systems.

13. Includes all equipment and subsystems not otherwise identified (examples: fire protection system, station maintenance equipment, fuel oil system, etc.).

14. Includes all final coal preparation equipment not included in account 312.1, the combustor, all coal injection equipment, and initial slag collection equipment at the combustor.

15. Include subaccounts for the nozzle, generator channel, and diffuser.

16. Include a subaccount for the magnet and each support system.

17. Includes all power conditioning and electrode control equipment.

18. Includes all equipment from the outside inlet through the delivery piping to the combustor; including the main air compressors, air compressor drives, low temperature air heaters, high temperature air heaters and hot gas piping.

19. Includes all seed unloading, storage, preparation, injection, transport, separation, and reprocessing subsystems and equipment.

20. Includes all other major support subsystems such as an oxygen system or a coal gasification system. Include a specific subaccount for each major support subsystem subdivided into its major components and subsystems.

21. Includes all equipment or subsystems provided specifically for research purposes (example: instrumented generator channel). Include a subaccount for each major research component or subsystem.

22. Includes all equipment or subsystems specifically provided to simulate equipment or subsystems which would normally be included in a commercial MHD power plant (example: heat rejection equipment used to simulate a steam bottoming plant). Include a subaccount for each major simulation component or subsystem.

23. Includes all transmission plant equipment located at the facility including the main transformers and switchyard.

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